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AFTERBURNER SYSTEMS STUDY

R. W. Rolke, et al

Shell Development Company

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Afterburner Systems Study

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E.R. Slater, T.T. Phillips, G.D. Towell

ENVIRONMENTAL PROTECTION AGENCY
Office of Air Programs
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The results are presented of a study of afterburner or fume incinerator technology for control of gaseous combustible emissions from stationary sources. The scope of the study included evaluation of current engineering technology, evaluation of existing afterburner systems, assessment of present practices and problems, determination of major sources and potential applications, and development of research recommendations. The main results of this study are presented as a handbook, allowing the potential user to be able to decide if his particular emission is amenable to afterburning and to obtain a rough estimate of cost and size of equipment needed. The user will also be made aware of potential problems and recommended design features. The user then would deal with the appropriate equipment supplier for details of equipment selection.

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Air Pollution Control Equipment
Afterburners
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Furnaces
Heat Recovery
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Shell Development Company
A Division of Shell Oil Company
Emeryville, California

Ib

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FORWARD

This study of afterburner technology was conducted under contract EHSD 71-3 for the Office of Air Programs of the Environmental Protection Agency with Dr. D. L. Harmon serving as project officer. The first part of this report presents a handbook of afterburner technology and the second part presents a review of the tasks performed, an inventory of major sources and potential afterburner applications, and research recommendations.

In addition to the authors, Messrs. C. M. Schlaudt, D. R. Hayes, G. A. Stenmark, J. S. Son and R. A. Brown participated in the project.

We acknowledge with thanks the cooperation and time of the staffs of the equipment manufacturers, afterburner users and the Air Pollution Control Districts, who provided much of the information upon which this study is based.

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Chapter 1. INTRODUCTION

A systems study of afterburner or fume incinerator technology for control of gaseous (or gasborne particulate) combustible emissions from stationary sources has been carried out. The scope of the study included evaluation of current engineering technology, evaluation of existing afterburner systems, assessment of present practices and problems, determination of major sources and potential applications, and development of research recommendations. Information was obtained from the published literature, equipment manufacturers, equipment users, air pollution control agencies, research institutes and universities. Visits were made to many of these sources of information in order to hold detailed technical discussions and observe equipment in operation. Questionnaires were sent to plants which were known to have afterburners in use or were likely to have them. The respondents provided information on general performance, costs and problems encountered with their units. Actual performance data were rarely available. Two main classes of afterburner are considered, thermal (direct flame) and catalytic. Disposal of liquids or solids by combustion are not included. Flares and afterburners which are part of a solids refuse disposal system are not included.

The current main uses of afterburners are for the control of odor and smoke with some applications in hydrocarbon and CO emission control. Afterburners are typically used to control low concentration combustible emissions. This comes about because of the safety requirement to keep the concentrations within the fume generating equipment below the lower flammability limit. Odor nuisance emissions are usually very low concentrations. A combustible gaseous emission produced at high concentrations can be used as a fuel. Alternatives to afterburning for gaseous emissions are adsorption on a solid bed or scrubbing with a solvent. Water scrubbing is generally not effective in controlling organic emissions. Alternatives to afterburning of organic particulate emissions are the use of a cyclone, filter, electrostatic precipitator or wet scrubber.

The main results of this study are presented as a handbook of afterburner technology. The potential user will be able to decide if his particular emission is amenable to afterburning and to obtain a rough estimate of cost and size of equipment that is needed. The potential user will also be made aware of potential problems and recommended design features. It is expected that the potential user would then deal with the appropriate equipment supplier for details of equipment selection. The handbook will be useful to equipment manufacturers and also to operators of afterburners for performance improvement. The handbook is contained in Chapters 1 through 20. The first part (Chapters 3 to 11) covers the description and performance of afterburner systems. The second part (Chapters 12 to 19) covers fundamentals and design principles. An annotated list of equipment suppliers is presented in Chapter 20. The remainder of the report contains documentation of the methods used in collecting the information, estimates of emissions and the applicability of afterburning to them, and recommended future R and D.

Chapter 2. SUMMARY

Afterburners or fume incinerators can be used to control the emission of gaseous or small particulate air pollutants which are combustible or thermally decompose at high temperatures. Hydrocarbons are oxidized to the products carbon dioxide and water. This oxidation may be carried out in thermal or catalytic afterburners, with the catalyst allowing use of a significantly lower operating temperature. Table 2-1 lists some of the processes from which emissions have been controlled using afterburners. Most applications have been for the abatement of smoke, odor, and (primarily in California) smog producing hydrocarbons. Effluent must be collected and ducted to the afterburner. In many cases, this costs more than the afterburner itself.

Table 2-1. APPLICATIONS OF AFTERBURNERS

Adhesive tape curing
Asphalt blowing
Brake lining ovens
Cat cracker regenerator off gas
Charcoal broilers
Coil and strip coating lines
Core ovens
Cupola furnace stacks
Deep fat frying
Fat rendering
Fiber glass curing
Herbicide and insecticide manufacturing off gas
Lithographing ovens
Meat smokehouses
Metal coating ovens
Metal reclaiming
Pulp and paper
Packing house effluents
Paint baking ovens
Paint removal facilities
Phthalic anhydride manufacturing off gas
Plastic curing ovens
Printing presses
Quench bath oil fumes
Resin and paint cooking
Roofing paper machine hoods
Rubber curing
Solvent degreasing
Solvent manufacturing off gas
Sulfur plant tail gas
Textile dryers
Varnish burn-off
Varnish kettles
Vinyl sponge curing
Wire enameling

It is important that a potential user work closely with an experienced manufacturer in order to obtain a satisfactory unit. The potential problems described in this report should be carefully considered in the design stage before equipment selection is made. The potential user can contribute significantly to the success of an afterburner installation by defining the concentration and flow rate ranges carefully. A low initial cost,minimal afterburner system is a poor investment if in a short time the unit falls apart or is incapable of meeting the target air pollution regulations. The information in the handbook should allow the prospective user to assess whether the system designs provided by various manufacturers avoid the potential problem areas.

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2.1 Thermal Afterburners

Thermal (direct-flame) afterburners destroy combustible pollutants through oxidation to CO_2 and water. Temperatures of 1400 - 1500°F are sufficient to obtain nearly complete conversion of most substances in .1 → .3 seconds residence time. Destruction of most hydrocarbons occurs rapidly at 1100 - 1200°F but oxidation of $CO \rightarrow CO_2$ requires the higher temperatures and residence time. Dense carbon smokes may require temperatures of up to 2000°F and/or longer residence times (~1 second), but liquid smokes are destroyed nearly as rapidly as gaseous hydrocarbons. (Gross liquid carryover into the afterburner can lead to problems, and should be prevented through use of knockout vessels.)

This handbook is mostly concerned with dilute fume incineration. Supplemental fuel must be burned to supply heat since the fume itself is not combustible and often contains only a few hundred parts per million of organics. Fuel to attain a 1400 - 1500°F afterburner temperature cannot be premixed with the entire fume stream since such a mixture is below the flammable limit. The fuel must be burned separately using ~50% of the fume for combustion air (if the fume has a high oxygen content and is non-fouling), and the hot combustion products (~2500°F) must be mixed with the remaining fume to achieve the ~1400°F temperature.

Mixing of bypassed fume and hot combustion gases is the most crucial step in attaining good afterburner performance. Typically, afterburners are designed with ~.5 second total residence time. This time is nearly all required for this mixing step and many designs fail to complete the mixing in the distance (time) available. Some fume escapes without being raised to a sufficiently high temperature. To meet a performance specification (if possible at all) more fuel must be burned than would be needed if mixing were complete. Distributed burners are placed directly in the fume stream and divide the flame into many individual jets or lines of flame surrounded by fume. This subdivision greatly speeds the mixing process, and these burners are well suited to use oxygen from the fume for combustion. The use of outside air requires an additional 30-50% of the fuel to be burned to heat it to 1400°F. Distributed burners are subject to fouling, have somewhat limited turndown, aren't available for use with oil fuel, may be difficult to use with outside air, and have a few other potential drawbacks. Therefore, many afterburners employ discrete burners which give either long or short point sources of flame. The mixing problem is much more difficult since there is no subdivision at the burner. Internal baffles are required in the relatively short afterburner chambers utilized in available designs.

Many designs stress "flame contact" in an attempt to mix fume and flame as rapidly as possible. This often leads to flame quenching and an increase in pollutants in the fume stream since as was mentioned above, complete fume/fuel mixing gives a noncombustible mixture. Fuel should be burned as rapidly as possible and the hot gases should be mixed with bypassed fume. Pressure drop must be expended in achieving good mixing through baffles and/or a long chamber. Mixing will be faster when there is initial fume/flame subdivision.

In comparison with boilers and industrial furnaces, thermal afterburners have low NO_x emission factors. The low operating temperatures and dilution of combustion products by fume results in stack NO_x effluent concentrations of only 20-30 ppm.

The presence of chlorine, sulfur, metals, nitrogen, etc., in the fume along with hydrocarbons does not rule out the use of thermal afterburners. These components may either pass through unchanged or be converted to the corresponding acid or oxide and will require additional flue gas treatment (e.g., wet scrubbing) if present in high enough concentration. Nitrogen compounds in the fume would be almost quantitavely converted to NO_x in the exhaust. Also, corrosion problems may arise and higher chamber temperatures may be required for complete destruction. Wherever unusual components will be present in the fume, sufficient temperature flexibility should be provided and/or prior testing should be done.

The normal configuration for construction of a thermal afterburner involves a steel outer shell, lined with a refractory material. The purpose of the refractory is to protect the steel shell from direct exposure to the effects of high temperatures and corrosive materials, and to improve thermal efficiency of the unit by limiting heat losses. The refractory serves as a thermal insulator, lowering the temperature from a maximum of more than 2000°F on the inside of the combustion chamber to a temperature of 250 - 400°F at the steel shell. At this temperature the shell retains most of its room temperature strength and is still hot enough to prevent condensation of water vapor on the surface, thereby reducing corrosive attack. Refractory is used in the form of bricks, castables, and dense boards made of pressed fibers.

Refractory structures are heavy, with densities running 45 lb/cu ft for the lightweight insulating firebrick and castable refractories up to 185 lb/cu ft for high alumina materials. Refractory wall thicknesses run typically from 4 to 8 inches. This weight adds considerably to the cost and difficulty of shipping, so that many afterburner manufacturers do not install the refractory until after the shell has been moved into place - expecially on the large, custom-designed units. The installation of full masonry construction calls for a relatively skilled mason. An unskilled mason can make mistakes that will lead to premature failure. Once in place the added weight of the masonry can be an important consideration in the structural design of the building, especially if the afterburner is to be installed on a roof or elevated platform. Because of its light weight, fibre-block wall construction is being used in some afterburners. However, this material has limited strength; care must be used in attaching it to the metal walls, and gas velocities must be kept low to avoid the possibility of erosive damage.

In order to avoid the heavy weight and slow heatup associated with refractory construction, some manufacturers have designed all metal units, employing incoming fume for cooling the outside of the combustion chamber walls. These units have been subject to many structural failures since metals have been used at the limit of their strength and oxidation resistance. Also thermal stresses are severe due to the large temperature gradients and the cool wall provides a path for fume escape without pollutant destruction.

It was previously mentioned that high concentration combustible gaseous emissions can be used as fuel to generate process heat. It may also be possible to use process heaters, utility furnaces, or steam boilers as afterburners for dilute fume streams and save the purchase cost and fuel costs of a separate afterburner. If such an existing furnace can provide sufficient residence time at a high enough temperature and the fume meets certain criteria listed in Chapter 5, fume can be added as part of the secondary air supply to the furnace. Uniform treatment of fume may be difficult to achieve so performance tests should be conducted.

There are many poorly designed afterburners and many of lightweight construction on the market. Many units are subject to structural failure since thermal expansion stresses can be severe and many are not capable of operation at 1400 - 1500°F. The "smoke eliminator" operating at 700 - 800°F is merely vaporizing liquid smoke droplets and producing lots of CO and aldehydes by quenching of the supplemental fuel flame fired across the stack.

The user survey produced 214 completed questionnaries on thermal afterburners. 89% of the units were regarded as satisfactory. The efficiency of hydrocarbon removal was reported for 66 units with two-thirds being in the 95-100% removal range and one-third in the 90-95% range. Generally units are adjusted during start up and a temperature found that results in adequate removal. No further testing is made and adequate performance is assumed if the same temperature is maintained. Maintenance on most units is not a major problem. The main operating problems involve safety controls, refractory linings, heat exchanger fouling or mechanical failure and bearing failure in fans.

2.2 Catalytic Afterburners

Catalytic afterburners are an alternative to thermal incinerators as a means for oxidizing gaseous, combustible contaminants to carbon dioxide and water. Their successful operation is limited to a more restricted range of applications than thermal afterburners; but where applicable, catalytic units offer the potentials of significantly lower fuel consumption and smaller, lighter-weight units. The basic elements of the catalytic unit are a preheat/mixing section, designed to achieve a uniformly preheated and distributed waste stream flow, and the catalyst bed or matrix, in which the major portion of the oxidation reactions take place.

Catalysts in use in afterburners are typically metal mesh-mats, ceramic honeycomb or other ceramic matrix elements with surface deposits of finely divided platinum metals. The oxidation of most hydrocarbons and CO occur rapidly in the range of 600 - 900°F over these catalysts (in contrast with 1200 - 1500°F required in thermal systems).

Catalytic systems are limited to applications in which the waste stream has a negligible particulate loading, and in which the combustible contaminant is gaseous or can be completely vaporized in the preheat section of the unit. Oxidation of methane presents a difficult problem for catalytic units and requires higher than normal operating temperature.

In use, the oxidation activity of the catalyst decreases. This decline in activity is accelerated by high temperature exposure, by particulates from the waste stream coating the catalyst surface (inhibiting contact of catalyst and waste stream), and by actual poisoning of the active catalytic sites by specific contaminants in the waste stream. Performance of the unit is maintained during catalyst deactivation by, (i) initial over-design of the catalyst bed, (ii) raising the operating temperature, (iii) periodic cleaning of catalyst, and (iv) eventual replacement of catalyst. Prediction of the frequency at which catalyst must be replaced is uncertain for new use. However, for some applications, catalytic afterburners have been in use for twenty years and manufacturers are knowledgeable concerning factors limiting catalyst life in these uses. In some cases catalyst has never been replaced whereas in others replacement is made every year.

The user survey only produced 24 completed questionnaires on catalytic afterburners. 30% of the units were regarded as satisfactory. Efficiency data was only reported in one case. A major maintenance problem was catalyst poisoning, reported in 17 units.

Because of lower operating temperature, enclosures for catalytic systems need not be so ruggedly constructed as for thermal systems. However, refractory or stainless steels are recommended for interior surfaces and parts exposed to preheat or afterburner temperatures. Configurations in current use are relatively simple, usually in-line placement of preheat burner and catalyst element. Design of the chamber and mounting of the catalyst element are based on obtaining uniform flow of preheated stream through the catalyst and avoiding bypassing of any part of this stream around the catalyst. Where possible, distributed burners are used to aid mixing of waste with preheat burner products.

Catalytic afterburning differs from many other industrial applications of heterogeneous catalysis in that afterburners are designed for unusually high rates of reactions (per unit volume of catalyst bed). Under such conditions, the overall rate of the catalyzed process becomes dependent on the rate of the mass transfer of combustible from the waste gas stream to the catalyst surface as well as on the rates of "chemical" processes taking place within the porous catalyst itself. An analysis which allows treating mass transfer and chemical contributions to determination of the rate, has been carried out for the catalytic afterburning process. A method for calculating oxidation performance based on this analysis is presented in Chapter 17. Correlations for mass transfer and heat transfer coefficients and friction factors to be used in these performance calculations have been developed for all of the catalyst matrix configurations in current use. Publicly available data on the rates of the "chemical" oxidation processes with these catalysts are scanty. A few estimated "chemical" rate parameters are included in this report.

The NO_x emissions from catalytic units will be low, typically 15 ppm, because of the low level of preheat.

2.3 Heat Recovery

Since the discharge of flue gases at a temperature level of 1300 - 1500°F to the stack represents a continual wastage of expensive heat energy, the afterburner user has a strong incentive to consider the many possible ways for heat recovery. Recovery methods include heat exchange between hot flue gases and incoming cool fume stream, recycling a fraction of the hot flue gases to the oven or process from which the fumes have arisen, and the use of the heat in other processing or heating loads, such as in generating steam for plant or process heating, for power generation, etc. The cost of the heat recovery equipment can probably be recovered in fuel savings in a few years,at least for fume flow rates greater than 5000 SCFM.

Applications of heat recovery methods to afterburners have often had problems.

1) Recovery must be feasible. Dirty streams may make exchangers inoperable, and plant heat balance may preclude effective use in ways other than exchangers.

2) Safety must be assured. Preheating the fume stream to a high temperature when it contains combustible components could initiate a fire.

3) The heat recovery equipment must be dependable, so that it does not cause inadvertant outages in the plant.

4) Payouts must be evaluated in terms of reasonable capital and operating costs, including realistic maintenance allowances.

5) There must be space available, or structural strength to carry it, if the heat recovery equipment is to be mounted near the afterburner or on the roof.

6) Deterioration of the heat recovery equipment must not jeopardize the pollution control function of the afterburner.

The most common type of heat exchange equipment is the cross flow tubular heat exchanger; in Chapter 20 equations, data, and procedures have been included for an approximate sizing of this type of exchanger. Rotary regenerative and switched packed-bed type exchangers may also be used, although they have not been applied widely.

Particular problems with heat exchange equipment relate first to the fouling action of contaminants present in the fume stream, and to the loss of performance and the fire hazards this may engender; secondly, to thermal expansion caused by unequal heating throughout the exchanger structure.

2.4 Auxiliary Equipment - Fume Generating Process

The afterburner system is centered around the afterburner itself and its heat exchangers, but consideration must also be given to the fume source, to process control systems, and to bringing fumes to the afterburner and discharging clean flue gas. Large fluctuations in fume flow rate and concentration must be limited if satisfactory performance is to be attained. Also, de-entrainment devices should be used to eliminate heavy carryover of liquid and inert particulates.

Failure of control instruments causing major or nuisance shutdowns is one of the most common complaints from the operators of fume incineration equipment. In other cases improperly installed temperature sensing elements or inadequate control systems have resulted in excessive fuel consumption, large temperature variations, and operation at temperatures too low to give acceptable pollutant destruction. Also, proper use of process controls will enable the user to operate drying ovens and similar equipment so as to obtain higher solvent levels in the fume stream, thereby reducing fuel costs. For these and other reasons, it is important that an afterburner system be chosen which includes effective and reliable process controls. Chapter 9 is intended to guide the selection of control and safety systems for afterburners. Descriptions of the control equipment generally available and typical control systems are given. Included are sections on combustion safeguards, temperature controls, and vapor concentration controls.

Fume collection and delivery to the afterburner is a very important part of the total system. Careful attention should be paid to its design since both capital and operating costs are sensitive to the manner in which it is handled. Long duct runs can cost more than the afterburner itself and condensation of combustibles can occur even in insulated ducts, causing a fire hazard. Proper hooding is crucial in order to insure that all fumes are collected and that air dilution is kept to a minimum. Once the fume is below 25% of the LEL, all extra air represents wasted fuel in the afterburner. Existing ventilation systems will probably require modification to reduce unnecessary dilution and to vent clean air directly to the atmosphere rather than through the afterburner.

Unless the fume generating process is operated at high pressure, the afterburner system should include its own blower to overcome the pressure drop (to provide mixing energy) through the unit. Blowers are normally placed upstream of the afterburner so they handle cool fume at the lowest possible volume. For a given pressure increase through the blower, the power consumption is proportional to the actual volume handled. Induced draft blowers are placed downstream of the afterburner so as to operate on clean gases and avoid fouling problems. Also the afterburner will then be under slight vacuum and any leaks will be inward. Fans are normally limited to operating temperatures below 700°F so heat exchangers and/or cooling air must be used to reduce the temperature of the hot exhaust gases. Induced draft blowers must be larger and consume more power. The volume at 600°F is twice that at room temperature, and dilution air also increases the volume.

2.5 Costs

Purchase, installed and operating costs of thermal and catalytic afterburners are covered in Chapter 11. The information is based on extensive discussions with manufacturers and users and also on replies by users to our questionnaires. Data have been presented graphically where possible.

Inevitably there is a wide range in costs depending on the afterburner type, the particular design and the nature of the installation. Thermal units without heat exchangers range from $1.50 to $4.00 per SCFM uninstalled depending on the type and on the size of the unit (larger units being cheaper per SCFM). Corresponding units with heat recovery are of the order of 3-1/2 to 5-1/2 times these costs. Catalytic units are generally somewhat higher in capital cost than thermal units with equivalent capacity. The average installation cost of a unit is of the same order as the purchase cost, but is highly dependent on the particular situation.

Operating costs for thermal units are largely a function of fuel requirements. Little operating manpower and maintenance is required. An average fuel cost for a thermal unit would be ~$4.50 per SCFM per year without heat exchange, and ~$2.50/SCFM/year with heat exchange. (Based on 24-hour per day operation, afterburner at 1400°F, no heating value in waste stream, waste stream temperature 400°F, gas 50¢/MSCF.)

Very few questionnaire returns were received from users of catalytic afterburners, so that the information is limited on these units. Operating costs are largely dependent on catalyst life, and fuel costs are not so important because of the lower operating temperature. Where the waste stream is clean, and a catalyst life in excess of two years can be counted on, this type of unit should be economically attractive.

2.6 Measuring Afterburner Performance

In order to assess performance of an afterburner, concentrations of critical components at the entrance and exit of the afterburner must be determined. Typically these are total hydrocarbons and carbon monoxide but may include specific hydrocarbons, halogens, particulates, etc. As a bare minimum the afterburner will have to meet the standards imposed by the local air pollution control agency, either emission standards for specific pollutants or a percentage destruction in passing through the afterburner. Other standards may prohibit any odor or visible plume. The operator of the fume incinerator will often be responsible for demonstrating compliance with regulations, either by doing his own testing or by hiring an outside laboratory. If the afterburner does not provide the pollutant destruction required, some investigation into interior performance of the unit must be made. Of primary concern are measurement of temperatures and flow patterns.

Certain measurements are also required before an afterburner is purchased. It is important that fume stream flowrate and composition be determined, both average values and the maximum variation. The amount and nature of entrained material will also influence design of the unit for reasons previously discussed.

In Chapter 10 an attempt is made to indicate the state of the art pertaining to the determination of a variety of air pollutants, measurement of temperature and determining flow patterns. Problems encountered in sampling and calibration are discussed. The methods available for determination of individual compounds or classes are summarized, and available instruments are listed.

2.7 Major Sources of Combustible Pollutants and Potential for Afterburner Control

The rate at which organic pollutants are emitted in the United States from both mobile and stationary sources has been estimated for 1968 and 1970, and projected through 1980, 1990, and 2000. The results are summarized in Table 21.1. The contribution of stationary sources is projected to rise from 50% of the total in 1970 to 85% in 2000. The projections of emissions from stationary sources were made with the assumption that the present levels of legislated emissions control would not change during the projection period. The projected mobile source emissions reflect the stringent automobile emission regulations which are scheduled to take effect in 1975.

Several of the largest sources shown in Table 21.1 are either mobile or are not amenable to afterburner control. Examples are motor vehicles, refuse disposal, gasoline marketing, etc. Conversely, many sources which do not emit nationally significant tonnages of combustible emissions are, non-the-less, prime candidates for afterburner control because they constitute severe local nuisances. Odor and smoke sources, e.g. rendering plants, coffee roasters, brake shoe debonding ovens, etc., frequently fall into this category. No attempt was made to estimate the total rates of emissions of effluents which are objectionable primarily because of malodor. Instead, a count of the number of point odor sources was used to give a measure of the severity of odor pollution problem. The results are shown in Table 21.2.

Emissions from sources amenable to afterburning have been characterized by the type of compound emitted. Each set of emissions has been assigned a ranking based upon toxicity, malodor, and photoreactivity to allow an importance comparison of the various pollutant sources. In addition, each source has been assigned a ranking based upon the relative ease of control by afterburning. The source characterization and rankings are given in Table 21.3.

Criteria to be considered when choosing between afterburners and other control devices for particular applications have been outlined. The costs associated with extensive afterburner application to control stationary sources emitting nationally significant amounts of combustible pollutants excluding, however, control of sources objectionable because of odor, is estimated to be $360 million for equipment purchase and installation. The fuel cost associated with this level of control (again excluding control of odorous sources) is estimated to be $452 million for the purchase of 0.931 trillion scf of natural gas annually. These estimates are based upon achieving the maximum feasible level of control which would reduce the total emissions shown in Table 21.1 for 1970 by 11.1 billion lbs per year, or about 15%. At this level of control, approximately half of the total emissions from the stationary sources suitable for afterburning would be incinerated.

It should be emphasized again that these costs make no provision for controlling sources objectionable primarily because of malodor or carbon monoxide, and they would be increased if these costs were included.

2.8 Proposed Research and Development Programs

Five-year programs at the $1 MM and $3 MM levels have been developed to provide technology where deficiencies exist, to generate the data required to predict and improve process performance and economics and extend the application of afterburners to additional air pollution sources.

Since little quantitative performance data was found in this study, field testing of thermal and catalytic units is recommended. Suitable analytical methods, preferably of an instrumental type, need to be selected and utilized.

Basic studies of thermal and catalytic oxidation kinetics at temperatures of interest are recommended to provide a basis for better design methods. Initial studies would be with hydrocarbons and would continue into odors and particulates. Final plans for the basic studies would consider heteroatoms (e.g., S, Cl, etc.). The poisoning and deactivation rates of catalyst are also needed. Alternative lower cost base metal catalyst studies are also recommended.

A substantial program is recommended in combustion design for thermal systems. A research combustor is proposed with provision for changing geometry readily and measuring internal temperatures, flow patterns and concentrations. The interaction of mixing and combustion needs careful study to advance the design techniques. Burner studies to find ways to get a large turndown and thereby stable operation with varying feeds is also recommended. More reliable design procedures can then be developed on this data base. Feeds studied in this research incinerator would be hydrocarbons, odors, particulates and compounds containing heteroatoms. Field testing of new designs are proposed to implement the research findings.

Chapter 3. THERMAL (DIRECT-FLAME) AFTERBURNERS: DESCRIPTION AND PERFORMANCE

Thermal afterburners are the most widely employed method to control the release of hydrocarbon fumes, especially solvents from coating processes and smokes. Pollutants are heated in the presence of oxygen to a temperature sufficient to allow complete oxidation in the residence time available. Typically the thermal afterburner consists of a refractory lined combustion chamber with a raw gas burner fired at one end to supply heat. Cold, oxygen-containing fume enters through the burner and acts as the air supply for the natural gas flame.

As is discussed in this Chapter, a thermal fume incinerator can be designed and operated to achieve essentially complete conversion of hydrocarbon pollutants to CO_2 and water. Its major drawbacks are the relatively high operating expense caused by the need to burn supplementary fuel (usually natural gas) to heat cold fume to the required high temperatures 1400° - 1500°F and the significant initial cost for a reliable unit. (Chapter 11 is a guide to estimating capital and operating costs for a particular application.) Heat recovery (Chapter 6) can be employed with an increase in first cost, but often with fuel savings large enough to rapidly return this added capital.

Catalytic afterburners are another way to reduce fuel requirements since in most cases pollutant oxidation will occur at a significantly lower temperature in the presence of catalyst. Successful operation requires fume streams which won't foul the catalyst and careful monitoring of the unit to insure that the catalyst has not lost its activity. As a result, in the majority of applications thermal fume incinerators are chosen for their promise of more trouble free operation, and heat recovery is employed to obtain comparable fuel savings. Sections 4.1.2 and 11.3 should be consulted for a more thorough comparison of thermal and catalytic afterburners.

This Chapter is designed to guide the selection of a thermal afterburner from among the many designs available and in use today. We first set forth the general requirements for satisfactory performance of any thermal afterburner. Then various design features are discussed as to how well (or poorly) they meet these requirements and how well they are expected to stand up mechanically. The material contained later in this handbook, especially Chapters 12 - 15, should be consulted for a discussion of the bases for the guidelines developed in this Chapter.

3.1 General Requirements for Satisfactory Performance

Successful fume incineration simply requires contacting pollutant molecules with sufficient oxygen at a high enough temperature for the oxidation reactions to go to completion in the time available at the high temperature. Nearly every previous discussion of incinerator design requirements indicates that the secret to success lies in the "3T's," time, temperature, and turbulence. Essenhigh, however, has commented[3-8]: "The chant of Time, Temperature, and Turbulence as being a useful basis for design is something of a mystical incantation that at best is a truism and at worst is misleading, if not actually wrong."

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The 3T's embody the correct notion of what is required for successful incineration, but fail to indicate how this can be put into practice. Especially troublesome is "turbulence."

Most fumes treated by afterburners are dilute, i.e., they do not contain sufficient combustible to support combustion. Therefore, supplementary fuel must be burned to generate heat, and the resulting hot combustion products must be mixed with the contaminated fume stream in order to raise the entire stream to a sufficiently high temperature. If the fume stream is not dilute and can support combustion without supplemental fuel, it can be treated as a fuel quality stream, either premixed or oxygen free, and burned in a normal waste gas burner. The resulting combustion temperature, >2200°F is sufficient to destroy any organic pollutant. In most cases the heating value can be recovered by using the concentrated fume stream as fuel in a boiler or process heater.

In most applications for afterburners, the fume stream is essentially contaminated air, and has adequate oxygen content (usually 15-21% m) both for burning the necessary preheating fuel and for oxidizing the contaminant. In a few potential applications, the contaminated waste stream is an inert gas (e.g. N_2) with a low hydrocarbon content. In such cases, air must be blended with the waste stream to supply oxygen. Some equipment manufacturers recommend that sufficient additional air be added to assure an oxygen content in the afterburner effluent of at least 4% m. This concentration allows oxidation of the contaminant to take place, with a satisfactory level of conversion, under temperature and residence time conditions within the range normally used in thermal afterburners.

3.1.1 Steps Involved in Dilute Fume Incineration

Figure 3-1 schematically indicates the sequence of steps which must occur if fume incineration is to achieve satisfactory destruction of pollutants in a dilute stream. In an operating afterburner these steps are not likely to be obvious since some of the equipment components are involved in several different steps. All fume probably enters through the burner, which fires into a single combustion chamber. As shown in the figure, at least part of a dilute fume stream must bypass the supplemental fuel flame. If intimate "flame contact" is attempted in a non-premixing system, serious flame quenching may occur. These points are often misunderstood and lead to design and performance deficiencies. Sections 3.1.3 and 3.1.4 expand on these crucial areas.

Once the supplemental fuel has been burned to yield its heating value, the cold fume and hot combustion products must be mixed to give a nearly uniform temperature to all fume flowing through the afterburner. This should be done as rapidly as possible without causing flame quenching so that sufficient residence time can be provided at the required temperature in a reasonably sized unit. As discussed in the Section 3.1.2 temperatures of 1400° - 1500° are normally adequate for complete destruction of pollutants if the stream is held for .1 - .3 seconds after the required temperature has been attained. This view of thermal afterburner operation ascribes no special role to "flame contact" which in fact is a beneficial concept as will be discussed in the Section 3.1.3.

Figure 3-1 follows

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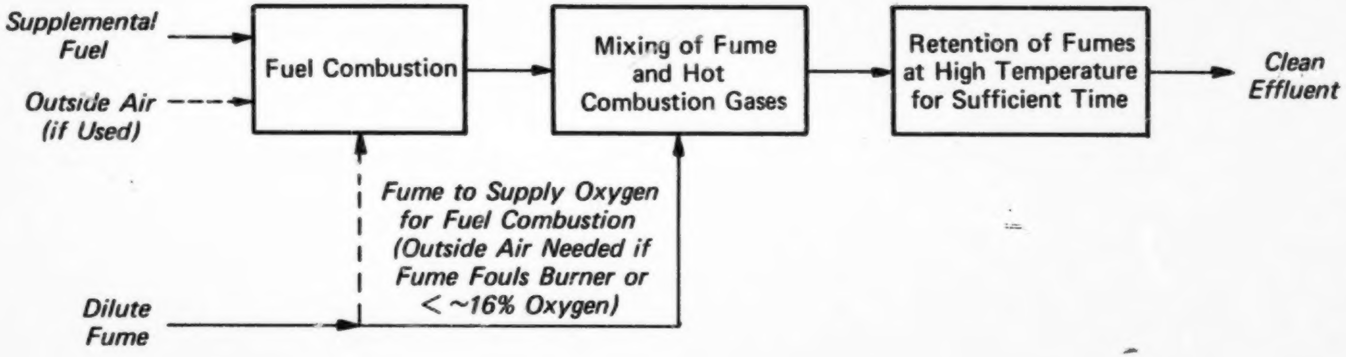
Supplemental Fuel
Outside Air (if Used)
Fuel Combustion
Mixing of Fume and Hot Combustion Gases
Retention of Fumes at High Temperature for Sufficient Time
Clean Effluent
Fume to Supply Oxygen for Fuel Combustion (Outside Air Needed if Fume Fouls Burner or < ~16% Oxygen)
Dilute Fume

Figure 3-1. STEPS REQUIRED FOR SUCCESSFUL INCINERATION OF DILUTE FUMES

For this reason, however, we refer to the broad class of non-catalytic afterburners as "thermal afterburners," whether or not a flame is utilized as the source of heat. The terms "thermal," "flame," and "direct-flame" afterburner are used interchangeably.

The greatest variation among different afterburner designs is in how well they achieve the goal of raising all of the fume to the required temperature for the required time. Most cases of poor performance are due to non-uniform temperatures and flows which allow some of the pollutants to escape without adequate treatment. This will be discussed in more detail as part of the consideration of various afterburner designs currently available. The consequences of the escape of some pollutants will depend on the situation. With extremely odorous components a few parts per million or less will give a smelly exhaust, so 99.9% destruction may be needed. Most hydrocarbon emission regulations have been written to require >90% destruction of pollutants in the afterburner. As is discussed in Sections 3.1.2.2 and 12.2.3 destruction rates are proportional to the pollutant concentrations. This means that the outlet concentration must be reduced to very low levels indeed if this inlet concentration is already low. For these reasons and reasons of fuel economy, it is important that unnecessary dilution be avoided. (Chapter 8 contains a discussion of proper fume-hood and ducting design to minimize dilution in excess of that required for safe operation.)

3.1.2 Temperature and Residence Time Requirements

Temperature and residence time requirements are discussed together since they are interchangeable to some degree - a higher temperature allows use of a shorter residence time and vice versa. Additional residence time involves a bigger combustion chamber and therefore a higher capital cost. However, additional volume is relatively cheap (see Chapter 11) and the residence time could be doubled for a 20 - 30% increase in capital. Operating temperature on the other hand bears a direct relationship to fuel usage, which is the major operating cost. Depending on specific considerations (fuel cost, hours/year operation, available space, etc.) it would often appear that temperature should be reduced and residence time increased.

However, this interchangeability of temperature and time is not of great practical significance since oxidation rates are very strongly temperature dependent. Figure 3-2a schematically indicates the general effects of temperature and residence time on oxidation rates in a flow through reactor. Over a narrow temperature range the rate increases from essentially zero to rates measured in milliseconds or less. At high temperatures complete conversion is controlled more by concentrations of pollutant and oxidant than by the temperature dependent rate. Since oxygen concentrations are almost always greater than 5% whereas pollutant concentration is reduced to a few parts per million, apparent rates are essentially independent of oxygen concentration and directly proportional (first order) to pollutant concentration.

The "ignition temperature" is used to characterize this marked change in extent of pollutant oxidation. However, as discussed in Section 12.1.1, ignition temperature is highly dependent on the conditions under which it is measured. Thus, autoignition temperatures (the temperature above which a combustible mixture of the substance and air must be raised to initiate combustion in the absence of a spark or flame) reported for many hydrocarbons have a large range. Among the widest ranges and highest temperatures are for toluene, 970°F - 1490°F and methane 1000°F - 1300°F. Also, as explained in Section 12.1.1, autoignition temperatures do not apply to hydrocarbon-air mixtures below the lower combustible limit (See Section 12.1.2.) For safety reasons, most afterburner applications involve dilute fume streams ranging from traces of combustibles up to 25% of the LEL. In order to achieve reasonable oxidation rates, temperatures 100°F or more above commonly reported ignition temperatures are usually required.

Table 3-1 shows typical ranges of residence times and operating temperatures for each of the major pollution abatement categories for which thermal afterburners are applicable. The range of conditions shown for each category is relatively wide. To some extent this range results from differences in oxidation rates of specific pollutants, due to their physical or chemical characteristics. However, the major factor contributing to the width of the range of conditions is the variability among afterburner designs in the effectiveness with which they mix fume with combustion products, and avoid by-passing or short circuiting of any significant part of the fume stream around the hot reaction zone. Those designs which most successfully utilize the residence time for oxidation of all of the fume at the average effluent temperature will permit attaining the conversion (or abatement) goal at an operating temperature and residence time near the lower end of the ranges quoted in Table 3-1. Those which are less successful will require higher temperature and longer residence time for the same abatement result.

3.1.2.1 Calculation of Residence Time

Total afterburner residence time is simply the total combustion chamber volume divided by the volumetric flow rate of fume and supplemental fuel combustion products. The volume of hot gases is directly related to the absolute temperature of the gas so the volumetric flow rate must be based on the actual volume at the combustion chamber temperature rather than the volume at standard conditions. Operation at 1400°F involves volume expansion ~3.5 times the volume at 60°F. Total average residence time, for cases where all oxygen for combustion is obtained from the fume can be estimated from

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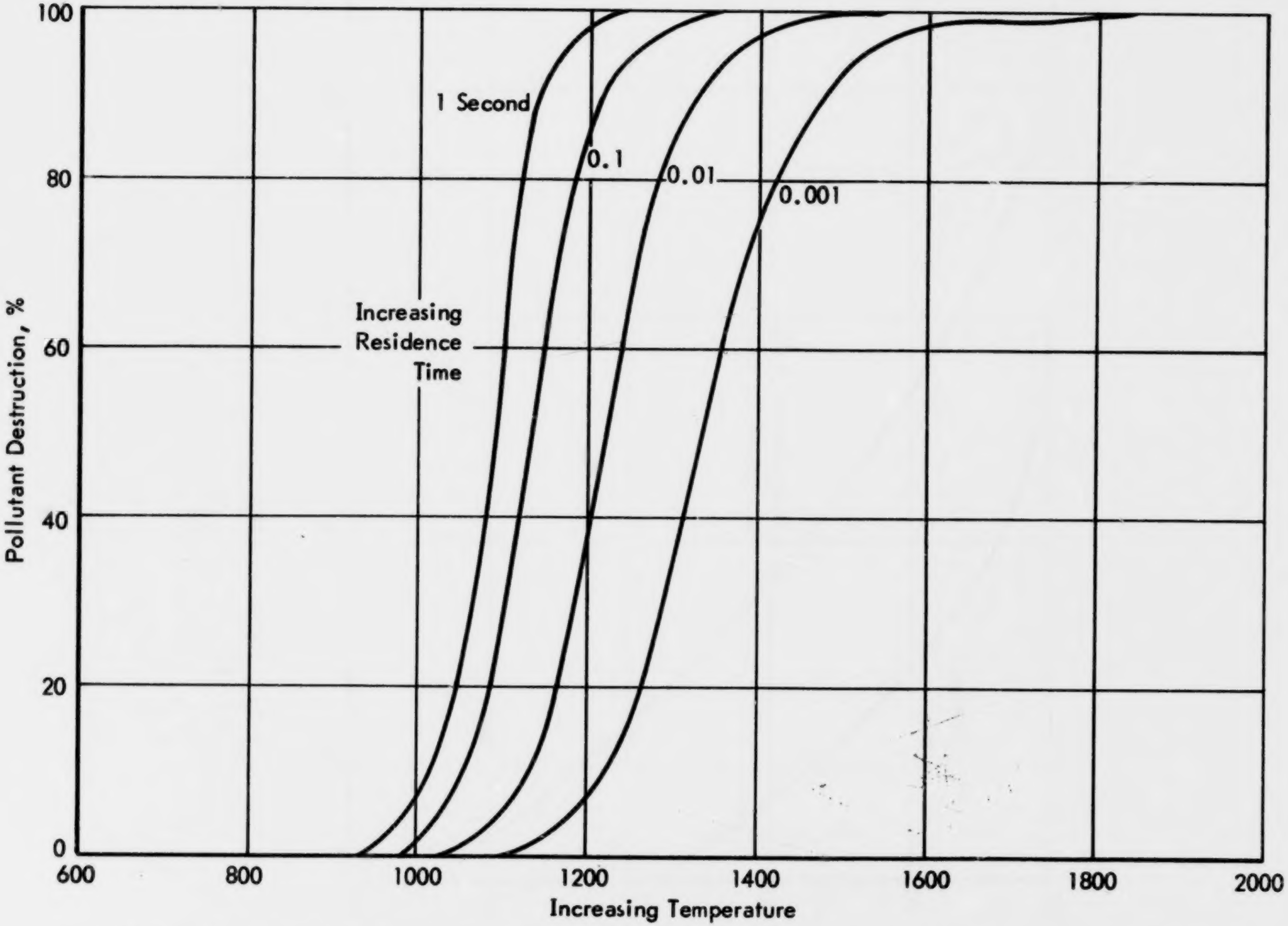

Figure 3-2a. COUPLED EFFECTS OF TEMPERATURE AND TIME ON RATE OF POLLUTANT OXIDATION

18e

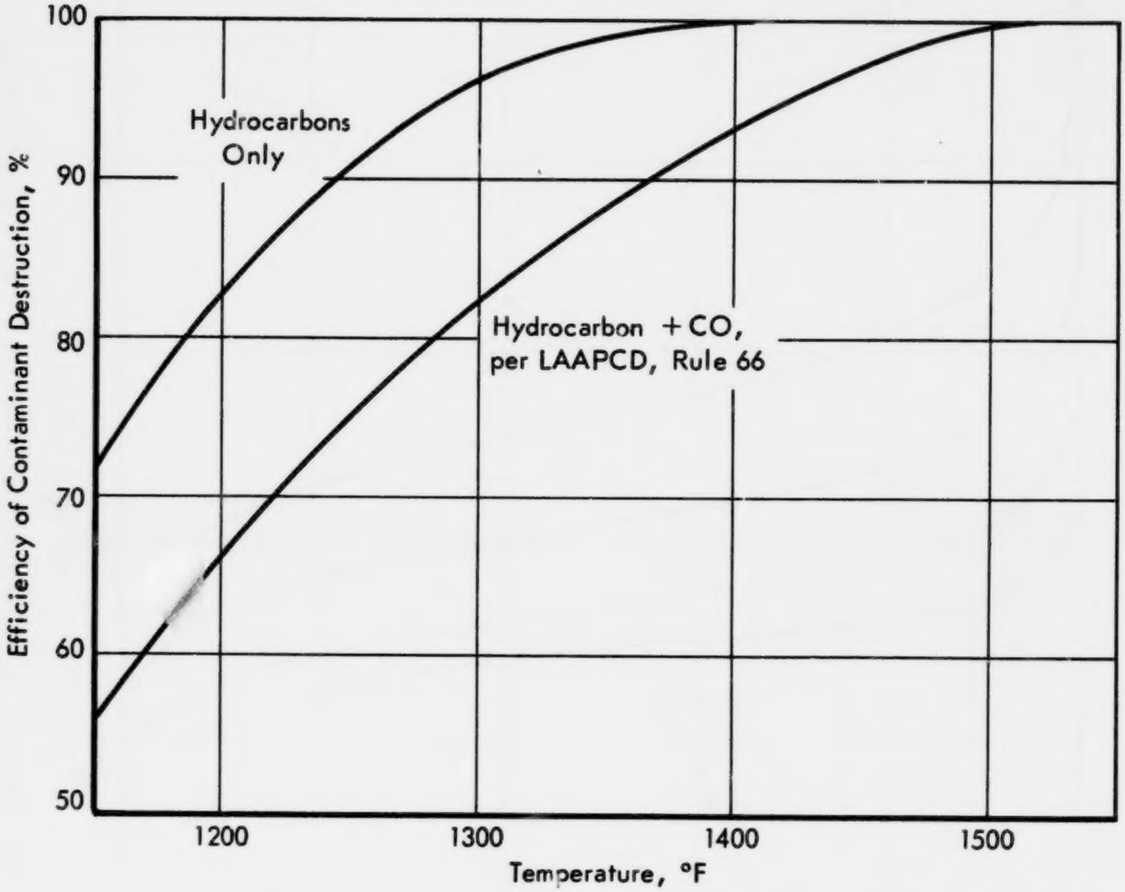

Figure 3-2b. TYPICAL EFFECT OF OPERATING TEMPERATURE ON EFFECTIVENESS OF THERMAL AFTERBURNER FOR DESTRUCTION OF HYDROCARBONS AND CO

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Table 3-1. THERMAL-AFTERBURNERS

Conditions Required for Satisfactory Performance in Various Abatement Applications

Abatement Category	Afterburner Residence Time (Sec)	Temperature (°F)
Hydrocarbon Emissions (90% + Destruction of HC)	0.3-0.5	1100-1250[a)]
Hydrocarbons + CO (90% + Destruction of HC + CO, as in LAAPCD Rule 66)	0.3-0.5	1250-1500
Odor		
(50-90% Destruction)	0.3-0.5	1000-1200
(90-99% Destruction)	0.3-0.5	1100-1300
(99% + Destruction)	0.3-0.5	1200-1500
Smokes and Plumes White Smoke (Liquid Mist)		
(Plume Abatement)	0.3-0.5	800-1000[b)]
(90% + Destruction of HC + CO)	0.3-0.5	1250-1500
Black Smoke (Soot and Combustible Particulates)	0.7-1.0	1400-2000

a) Temperatures of 1400-1500°F may be required if the hydrocarbon has a significant content of any of the following: methane, cellosolve, substituted aromatics (e.g. toluene, xylenes).

b) Operation for plume abatement only is not recommended, since this merely converts a visible hydrocarbon emission to an invisible one, and frequently creates a new odor problem due to partial oxidation in the afterburner.

$$\tau = \frac{V \cdot 60}{(SCFM)3.5} \qquad (3\text{-}1)$$

where τ = total average residence time, seconds

SCFM = fume flowrate to afterburner at 60°F and atmospheric pressure (standard conditions), ft^3/minute

V = combustion chamber volume (following burner and prior to heat exchangers, dilution air, etc.), ft^3

For more precise calculations the factor 3.5 should be replaced by $(T(°F)_{\text{combustion chamber}} + 460)/520$ and (SCFM) should include supplemental fuel and any outside air added through the burner. When outside air is not used, methane will contribute less than 3% to the volume. However, if the required supplemental fuel is burned using only outside air the volume may be increased by 50% or more.

Our survey showed that afterburners are operating with total average residence times which vary from 0.1 to 5 seconds. It should be stressed that this is the total residence time and is not equivalent to the time during which fume is held at the required temperature. A significant amount of time (often greater than the total residence time) is required to bring the cold fume up to temperature. Also, all elements of fume do not spend an equal amount of time in the combustion chamber. Some are swept out very quickly while others are trapped for a much longer time. This variation in residence times is a function of flow patterns attained and can greatly affect performance of the unit.

3.1.2.2 Hydrocarbon Destruction

Destruction of most hydrocarbons occurs very rapidly at temperatures in excess of 1100° - 1200°F. Possible exceptions are methane, cellosolve, and benzene derivatives, like toluene, which are stable molecules and require a higher temperature (~1400°F) for oxidation to occur in a few tenths of a second. More information on expected rates of hydrocarbon destruction is contained in Section 12.3.1. Except for a few compounds, required conditions for destruction are not sensitive to the hydrocarbons involved. (Methane as noted is difficult to destroy at temperatures below 1400°F. However, in most cases the presence of methane in the fume indicates inefficient combustion in the burners in the drying ovens, i.e., flame quenching. Adjustment of these burners will normally markedly reduce inlet methane concentration to the afterburner.)

If hydrocarbons are found in the flue gas of an afterburner operating at a nominal combustion chamber temperature above 1400°F (or above 1200°F for all but a few hydrocarbons), it is due to poor mixing and non-uniform treatment of the fume stream rather than too low a temperature. (Excluding the possibility that the indicated temperature is higher than the actual temperature, either because of the need for instrument recalibration or because the thermocouple "sees" the high temperature flame.) A higher nominal temperature will usually reduce outlet hydrocarbons in this case but the benefit comes from raising the temperature of that portion of the fume which sees a much lower temperature or shorter residence time than the average, rather than through increased efficiency due to the increase in the average temperature. In effect, the extra fuel is wasted.

At these same temperatures most hydrocarbons should disappear in ~.1 second or less. Since most afterburners have total average residence times many times this long, slow mixing and bypassing must be blamed for any failure to achieve complete destruction. Increasing the residence time should prove beneficial since more time is available for mixing of hot combustion gases and cold fume. However, if mixing is very poor to start with, a doubling of residence time may produce little noticeable effect. Also, if residence time is increased by decreasing throughput, performance is likely to be poorer since the gases will be flowing at a lower velocity and will mix more slowly.

Hydrocarbon destruction rates will increase with initial fume concentration and better cleanup will be possible with a higher concentration given the same mixing efficiency and afterburner temperature. This is partly due to the first order dependence of oxidation rate on pollutant concentration and partly due to the significant local heat of reaction which is released during oxidation. But primarily the ability to achieve better cleanup is due to increased concentrations of free radicals which actually carry out the oxidation. (See Section 12.2.3 for additional discussion.) This improved performance may not be observed, if the higher heating value of the more concentrated fume, results in much poorer mixing because there is less mixing energy in the smaller supplemental fuel flames. However, this falloff of rate with concentration should be kept in mind when dealing with very odorous substances or the need to achieve a high destruction efficiency with a very dilute incoming fume.

3.1.2.3 Carbon Monoxide Cleanup

In the previous section it was mentioned that operating temperatures of 1400°F and above are rarely required to achieve hydrocarbon destruction except in the case of poor mixing of cold fume and hot combustion products and a few stable hydrocarbons. Such high temperatures do seem to be required, however, to obtain sufficient oxidation of $CO \rightarrow CO_2$ in <.5 seconds. As is discussed in Sections 12.2 and 12.3, the time required for the oxidation of CO is >10X that required for all of the steps involved in the oxidation of hydrocarbons to CO. Carbon monoxide cleanup is required since it too is a pollutant which is detrimental to health and is thought by some to lead to photochemical smog. Los Angeles Rule 66 includes the net production of CO in calculating afterburner efficiency.

$$\text{Afterburner Efficiency} = \frac{\{\text{Hydrocarbons}\}_{IN} - \{\text{Hydrocarbons}\}_{OUT} + \{CO\}_{IN} - \{CO\}_{OUT}}{\{\text{Hydrocarbons}\}_{IN}}$$

where { } indicates concentration expressed as C_1.
Thus the afterburner may achieve 98% destruction of hydrocarbons but fail to achieve the 90% minimum efficiency because of high CO production.

Section 12.3.2 contains some rate expressions for CO oxidation flames and also some estimates of the oxidation rate at concentrations and temperatures of interest in fume incineration. Afterburner experience shows that temperatures of 1400° - 1450°F are required with an actual residence time at this temperature (after mixing is nearly complete) of .2 - .4 seconds to achieve nearly complete oxidation of CO - CO_2 (less than ~200 PPM of CO in the flue gas) see Figure 3-2b. Units with poor mixing patterns exhibit outlet CO concentrations >1000 PPM though temperatures are >1400°F. A few units have provided sufficient CO cleanup at temperatures as low as 1200°F. However, this has normally been associated with a high solvent concentration, and as was mentioned previously, the higher the concentration, the easier pollutant destruction will be. Effective oxidation rates are higher for the reasons discussed in Section 12.2.3.

3.1.2.4 Odors, Aldehydes and Other Oxygenated Hydrocarbons

Afterburners are often used to destroy odorous fumes containing only 100 PPM or less of the odor producing substances. These are likely to be aldehydes, organic acids or sulfur containing compounds. Oxygenated hydrocarbons are intermediates in the oxidation of hydrocarbons, and as a result they are easier to destroy than the original hydrocarbon. Often afterburner temperatures of 1000°F or less are adequate for odor removal applications. Little if any CO will be oxidized at these temperatures in the time allowed, but so little is produced from >100 PPM of odorous material that it can be released without concern. However, care must be taken that the fuel necessary to heat the fume stream is burned in such a way that flame quenching does not occur. (See Section 3.1.3.) If it does, high concentrations of CO and methane from the fuel burner will be found in the stack gases.

It is especially important that good mixing patterns be established when dealing with extremely odorous compounds. Temperature uniformity is required across the combustion chamber and bypass must be avoided if the odor is to be completely destroyed. Regulations dealing with solvent emission control can be met even if 5 - 10% of the original hydrocarbon escapes treatment in the afterburner. However, odors may still be a problem if less than 1% of the original concentration remains in the flue gas.

3.1.2.5 Liquid Smokes and Droplets

Outside of California smoke and odor control is the most common application of afterburners and most of these are operating at 1000°F or below. As mentioned in the previous section, CO oxidation will not occur at an appreciable rate in this temperature range. Condensed hydrocarbon smokes are given off by many processes such as rubber curing, meat smoking, asphalt blowing, deep fat frying, etc. Much of the smoke given off from refuse incineration and improper oil burning is also liquid rather than dense carbon smoke.

Since organic liquids which are likely to be found in fume streams have boiling points below 600 - 700°F, the smoke plume can be made to disappear by merely heating the effluent gases to 800°F or less. About all that such low temperature afterburners do is to vaporize the visible smoke and render it invisible. That's all they would do if supplemental fuel to supply heat were burned externally in an efficient manner. However, a common arrangement is to place a burner at the base of the stack and fire directly into the smoke containing effluent gases. Gross flame quenching occurs and much CO, aldehydes, and unburned fuel is found in the "smoke free" stack gases. The aldehydes create a bad odor problem. In many cases there is less total pollution when the afterburner is not operating.

Liquid smokes require the same ~1400°F afterburner temperatures that gaseous fumes require. In Section 12.3.3 it is shown that the additional time required to burn a 100 micron hydrocarbon droplet is ~.02 seconds at 1400°F. Since drops bigger than 50 - 100 microns are easily removed in simple cyclones and knockout vessels, no special problems are introduced relative to rates of pollutant destruction. Certain practical problems such as fouling of burners and buildup in duct work are discussed elsewhere.

3.1.2.6 Soot and Combustible Particulates

As contrasted to liquid smokes, solid carbon (soot) smokes and other combustible particulates are expected to increase the required afterburner temperature and/or residence time. (See Section 12.3.4.) Particles larger than 50 - 100 microns should be removed ahead of the afterburner using simple cyclones or settling chambers. If particles are larger than 5 microns (5×10^{-4} centimeters), residence times of a full .5 seconds may be needed at 1400° - 1500°F to achieve complete burnup. Since mixing and particle heating require additional time, total average residence times of .7 - 1.0 seconds or higher temperature capabilities should be provided for large particulates. One method of providing longer residence time without increasing the combustion chamber size, is to use tangentially fired burners in a vertical upflow afterburner. (See Section 3.3.2.3.) The resulting cyclonic action and gravity will hold large particles near the bottom of the afterburner chamber until they burn up.

Small soot or other dense carbon particles present another problem. Soot typically has particle diameters <.1 micron but may agglomerate to several times this size. Soot will not be affected by any cyclonic forces and appears to exhibit surface burning rates two orders of magnitude less than these for pulverized coal or coal char. Extreme temperature and residence time requirements are calculated (Section 12.3.4) by extrapolating soot combustion data into the temperature regions of interest. Temperatures of up to 2000°F and residence times of 1 second or more may be required for dense carbon smokes.

Afterburner experience with wire burnoff, smelting operations, and cloth carbonization shows that temperatures of over 1700°F and residence times greater than .5 seconds are often needed to clean up the black smoke plume. The required severity of treatment seems to depend on the amount of volatile matter present and porosity of the particles. As mentioned in the previous section, most smoke from refuse incinerators is a liquid smoke and is easily burned up at temperatures of 1400° - 1500°F. However, smoke from plastics incineration sometimes requires temperatures greater than 1700°F for destruction in a secondary combustion chamber. Since limited information is available, some testing and/or operating flexibility (e.g., high temperature capability) is recommended where no direct experience with similar dense carbon smokes is available.

3.1.2.7 Nitrogen Oxides (NO_x) Formation

In afterburners, as in any combustion equipment where fuel is burned with air in a flame, some reaction of nitrogen and oxygen from the air occurs, forming some nitric oxide (NO) and nitrogen dioxide (NO_2). (For measurement and discussion, these two oxides of nitrogen are referred to cumulatively as NO_x.) Substantially all of the NO_x formed in a thermal afterburner is formed in the high temperature region (>2800°F) of the burner flame itself. At the temperature of the main residence-time section of the afterburner (1200-1500°F), the overall rate of reaction of nitrogen with oxygen is too slow for significant formation of NO_x. However, in occasional afterburner applications where nitrogen-containing compounds, e.g. NH_3, amines, pyridines, are present in the fume contaminants, additional NO_x may be formed during fume oxidation. This formation of NO_x from oxidation of nitrogen-containing compounds is discussed below in section 3.1.4.8.

The normal size of afterburners and the types of burners used to facilitate mixing of fume with combustion products, result in relatively small flame size, with rapid quenching of flame products from >3000°F to ~1500°F. Thus, nitrogen and oxygen passing through the flame region have a very short residence time in the high temperature region where temperatures are high enough for NO_x formation. In comparison with larger combustion equipment (e.g. boilers and other industrial furnaces), afterburners have relatively low NO_x "emission factors", (i.e. the amount of NO_x formed per unit of fuel burned).3-9) For gas-fired afterburners, the "emission factor" has been found to be typically 0.05-0.1 lb NO_x/MM Btu heating value of the fuel burned.3-9) (In emission factors, the "lb NO_x" is customarily expressed in terms of NO_2 -- as though all the NO_x were present as NO_2 with a molecular weight of 46.) No test results are available from oil-fired afterburners. However, from results of tests with other combustion equipment, oil-fired afterburners would be expected to have NO_x emission factors higher by a factor of 2-3 than gas-fired units.3-9)

The use of preheated combustion air (which reduces fuel consumption) leads to higher flame temperatures and more rapid NO_x formation. Similarly, preheated or heat-exchanged fume may lead to slower flame quenching and correspondingly higher NO_x formation per unit of fuel actually burned. Reference 3-9) recommends the use of the emission factor together with the effective heating value -- the heating value of the fuel actually burned plus the heating value (above ambient temperature) of the preheated stream -- in estimating NO_x emissions from furnaces with air preheaters. The extension of this concept to afterburners with heat exchanger will probably result in a slight overestimation of NO_x emissions. However, it appears better to err in this direction than to underestimate emissions by neglecting any effect of preheating on NO_x formation. Using the assumption of effective heating value and emission factors of 0.08 for gas-fired units and 0.18 for oil-fired units, the emission rate and concentration in afterburner effluents given in Table 3-2 have been estimated.

Table 3-2. ESTIMATED NO_x EMISSIONS FOR THERMAL AFTERBURNERS

	Operating Temperature (°F)	NO_x Emission Rate $\left(\frac{\text{lb } NO_x/\text{Hr}}{\text{MSCFM Fume}}\right)$	NO_x Concn in Effluent (ppm)
Gas fired afterburners			
Fume used as air supply for burner	1200 1500	0.13 0.16	18 22
External combustion Air for burner	1200 1500	0.18 0.22	18 22
Oil fired afterburners			
External combustion Air for burner	1200 1500	0.40 0.50	40 50

The afterburner design concept involves dilution of primary flame products by mixing with cold fume stream to achieve the afterburner operating temperature. Thus, the NO_x formed in the flame is emitted from the stack at only 0.2-0.3 times the concentration at which it was formed in the flame zone. In contrast, in efficient larger scale combustion equipment, operated with minimum excess air, combustion gases are cooled to stack temperature by heat exchange (with boiler tubes or other heat transfer surfaces) rather than by dilution and NO_x is emitted at the concentration at which it formed in the flame. Emissions from afterburners, already lower than from other combustion equipment, on an emission factor basis, are also at even lower concentrations because of the dilution effect. For gas fired afterburners, estimated concentrations of NO_x in the stack effluent of 18-22 ppm are shown in Table 3-2.

Experimental studies of NO_x emissions from afterburners were included in a 1960 study by the LACAPCD[3-9] of NO_x emissions from stationary sources. The report of that study characterized measurements on 24 afterburners by emission factors of 0.05-0.08 lb NO_x/MM Btu. One burner manufacturer reported measurements of NO_x concentrations of < 20 ppm when gas-fired distributed burners are operated for 1500°F or lower afterburning. Test data supplied by two afterburner users on three well-designed thermal afterburners operating on paint baking oven effluents indicated stack concentrations of NO_x of 20-35 ppm when these units were operated at temperatures of 1400-1500°F. Data[3-10] obtained by Scott Research (Plumsteadville, Pa.) on two thermal afterburners in tests made for the National Coil Coaters Association are summarized below.

	Afterburner Unit	
	A	B
Inlet volumetric flow (SCFM)	6900	3000
Outlet gas temp (°F)	1290	1070
Increase in NO (ppm)	25	8.4
Increase in NO_2 (ppm)	2.2	3.0
Emission factor (lb/MM Btu)	0.12	0.06

(These data indicate a distribution between the two oxides of nitrogen of 70-90% nitric oxide, 10-30% nitrogen dioxide.) All of these available experimental measurements are reasonably consistent both with each other, and with the estimation basis used to construct Table 3-2.

3.1.2.8 Wastes Containing Chlorine, Sulfur, Phosphorous, Nitrogen, Metals, or Other Heteroatoms

Afterburners are used for the control of combustible air pollutants but components other than carbon, hydrogen, and oxygen may be present. These can either be chemically bound with the combustible, as a chlorinated solvent, or merely be an inorganic dust carried along in the fume stream. In the latter case, one would not expect any change to occur at typical afterburner temperatures, and the dusts will appear in the flue gas. Chemically bound heteroatoms will form the corresponding oxide or acid, e.g., chlorine will form hydrochloric acid (or free chlorine), sulfur forms sulfur dioxide, metals form their oxides commonly called ash. Section 12.5.2 discusses the likely fate of most heteroactoms. If concentrations of these products are high enough, e.g., >200 PPM SO_2 or >50 PPM HCl, >.1 gr/SCF ash, the afterburner alone will not provide sufficient pollution control. Wet scrubbers, cyclones, filters, etc. must be considered for flue gas treatment.

The presence of various heteroatoms does not rule out the use of thermal afterburners as it would catalytic units because of catalyst poisoning (Section 4.2.4). In fact, many fume incinerators are used to burn poisonous H_2S to SO_2 and NO_2 to a mixture of NO and N_2. However, the heteroatoms, particularly the halogens, do inhibit the oxidation reactions (Section 12.3.5), and even at low concentration they may cause higher temperatures and longer times to be needed for complete destruction of pollutants. No firm guidelines can be provided so testing is recommended prior to design of an afterburner to handle such wastes. Corrosion problems are especially severe when acidic components are present in the flue gas. This must be kept in mind when choosing construction materials for the afterburner and associated duct work.

Nitrogen containing fume components (e.g. NH_3, amines, pyridines, etc.) are of particular interest, since their oxidation can contribute to the total NO_x emissions from an afterburner. No direct information is available on the extent to which organic nitrogen in the fume is oxidized to NO_x. However, chemically bound nitrogen in fuels oxidized in flames is converted to NO_x in yields ranging from 20-100% of stoichiometric.(3-11) The remainder, not converted to NO_x, is almost entirely converted to N_2. The yield of NO_x increases with decreasing concentration in the fuel -- trace quantities are converted quantitatively, while major concentrations show lower yields. The concentrations of contaminants in fume streams approach most closely the situation corresponding to low or trace concentration of nitrogen compounds in fuels. It should be pointed out that the comparison between flame burning and fume oxidation is not ideal, since the two processes do take place at quite different temperatures. However, from what is known(3-12) of the mechanism of formation of NO_x in flames from reduced nitrogen compounds in the fuel, it is expected that the lower temperature of fume afterburning would increase selectivity to NO_x (over N_2). Thus, one would expect in most situations to see nearly quantitative yields of NO_x from nitrogen compounds present in the fume stream.

3.1.3 The Role of Flame Contact - Dilute Fumes Must ByPass

Many afterburner designers feel that contact between the supplemental fuel flame and the fume is needed for satisfactory pollutant destruction. Since this may lead to improved mixing between cold fume and hot combustion products and excess free radicals from the flame will enhance oxidation rates, the concept of "flame contact" is a good one. However, all of the fume cannot be put through the flame since a non-combustible mixture would result when fuel rate is set for a 1400° - 1500°F net temperature. Therefore, too much zeal in pursuing the goal of flame contact will lead instead of flame quenching and possibly an increase in pollutants.

3.1.3.1 Fuel Requirements and Stream ByPassing

As is discussed in Section 12.1.2, combustible mixtures are those which release enough heat through the oxidation reactions to raise the neighboring unburned fuel and air above their "ignition temperature" and overcome heat losses. The lowest concentration at which this will occur is the lower combustion (explosive) limit, the LEL. This lowest combustible mixture yields a flame temperature ~2300° - 2400°F for most hydrocarbon substances burned in air. (Table 12-1) (Methane burns as low as ~2100°F.) If the mixture inlet temperature is raised, less combustible is required since the minimum flame temperature capable of sustaining combustion is more or less fixed.

Afterburner fuel requirements are chosen to provide the minimum required temperature for pollutant destruction in the time available at this temperature. As discussed in Section 3.1.4 temperatures above 1400° - 1500°F are normally not required. If this fuel is premixed with the cold fume a noncombustible mixture will result. The LEL for methane in air at room temperature is 5%v compared to the less than 3% methane required to heat a cold, very dilute fume to 1400°F. Heat balance calculations are used to determine the heat required to raise fume and combustion products to a specified temperature and the fuel which must be burned to release this heat. Heat released during oxidation of the pollutant is often not negligible and must be included. (A 25% LEL fume stream will give a ~500° - 700° temperature increase during oxidation depending on the specific pollutant and temperature level.) Chapter 11 contains further discussion and useful graphs relating to the calculation of afterburner fuel requirements.

As fume temperature and/or fume concentration increases, a heat balance calculation predicts that no fuel is needed to achieve a 1400°F combustion chamber temperature. This occurs at ~50% LEL and no preheat or ~25% LEL and an 800°F preheat temperature. A certain amount of supplemental fuel is always required, however, to insure that incoming fume will be raised above its "ignition temperature" so that its heat content will be released. For most hydrocarbons 900° - 1000°F temperature rise is sufficient (Section 12.1.1) and fuel burners should be designed with enough capacity to raise cold fume by a minimum of 1000°F. The burner can then be "turned down" to a point where stable operation is insured. Streams more concentrated than ~50% LEL (fume in the range from 50% LEL to the upper combustible limit should not be encountered for safety reasons) should not be preheated and will require combustion chamber temperatures above 1400°F. Many afterburner failures have been caused by excessive combustion chamber temperatures resulting from a high combustibles content in the fume stream. This is often caused by entrainment of liquid droplets which were present due to poorly designed or nonexistent cyclones and knockout vessels.

Wherever possible, oxygen in the fume should be used to burn the supplemental fuel so part of the fume passes through the flame and part bypasses and is mixed with the combustion products downstream of the flame. (Use of outside air will increase fuel requirements by 30 - 50% and requires all of the fume to be mixed in downstream of the flame.) For a 1400°F combustion chamber temperature 48% of a cold (80°F) fume stream can be mixed with methane to give a 5.9%v mixture. (5%v is the LEL.) Thus, approximately half of the fume must bypass the flame and be mixed with the combustion products. If fume enters at 400°F instead (or equivalently has a 320°F heating value), less methane must be burned to provide a 1400°F chamber temperature and more fume must bypass (~59%) to maintain the same flame temperature (2500°F) as before. Thus good mixing downstream of the burner becomes more crucial for higher preheat temperatures and higher inlet concentrations.

3.1.3.2 Benefits of Flame Contact

As is discussed in Section 12.2.1, oxidation reactions in flames proceed very rapidly and primarily involve reactions of OH, O, and H free radicals with the fuel molecules and intermediate components. Oxidation rates are proportional to the concentrations of these radicals and radical concentrations are several orders of magnitude higher in a flame than they are in the absence of flame at 1400°F. As we have just noted, it is not possible to sustain a hydrocarbon/air flame at 1400°F, but the high radical concentrations take ~.01 - .03 seconds to decay after fuel has been consumed in the >2200°F flame. Therefore, pollutants in the fume will be destroyed more rapidly if hot combustion products and fume are rapidly mixed; i.e., "flame contact." Care is required, however, since CO oxidation to CO_2 in the flame occurs on the same time scale as the decay of excess free radicals so rapid mixing will lead to quenching. Also, as is discussed in Sections 3.1.2 and 12.3, required oxidation times for hydrocarbons at 1400°F in the absence of flame are much shorter than the ~.5 second typical afterburner residence time.

A more beneficial result of attempts to provide flame contact has been the development of distributed burners (Section 3.2). In these burners the supplemental fuel flame is spread out across the fume inlet to the combustion chamber and mixing of bypassed fume and hot combustion products is more easily accomplished because some subdivision has already occurred.

3.1.4 Quenching Must Be Avoided - Presence of Aldehydes and CO in the Flue Gas

Temperatures in the range of 1400° - 1500°F are required for reasonable rates of complete conversion of hydrocarbons to CO_2 and H_2O. In Section 12.2 we discuss the fact that oxidation reactions involve free radical chain carriers and that aldehydes, organic acids, and carbon monoxide are formed as intermediates. These intermediates and the original hydrocarbon will be found in the afterburner flue gases if quenching occurs and the oxidation reactions are essentially terminated.

Quenching is most likely to occur in local regions in the combustion chamber due to cold spots or too rapid mixing of fume with the flame. A few possible quenching situations are:

1. Fume which has been heated to a sufficiently high temperature may be cooled due to contact with a cold wall or mixing with cold fume.

2. If cold fume is rapidly mixed with the supplemental fuel flame (to achieve flame contact) some of the fuel and CO from the flame may be rapidly cooled causing reaction to stop.

3. Some fume or quenched gases from the flame may pass through the afterburner without reaching the nominal combustion chamber temperature. (Poor mixing or bypassing can cause large variations in temperature and residence time.) If only moderate temperatures are attained (~1000°F) aldehydes will be prevalent and will remain when the flue gases are quenched after leaving the combustion chamber. This is also a problem with plume eliminators designed to operate below 1000°F.

4. High surface area within the combustion chamber may serve as a third body for radical recombination reactions, resulting in partial quenching and a reduction in chain carrying free radical concentration and oxidation rate. (See Section 12.2)

Severe quenching by any of these mechanisms can lead to a worse pollution problem with a poorly designed afterburner in operation than when it is shut off.

3.2 Thermal Afterburner Design Features - Systems Utilizing Distributed Burners

It was noted in Section 3.1 that large fuel savings (up to 50% or more) are associated with using oxygen in the fume for combustion of supplemental fuel rather than bringing in outside air which must also be heated to 1400°F. It was also pointed out that ~1/2 of the fume stream must bypass the supplemental fuel flame and that achieving good mixing between this bypassed fume and the hot combustion products is the most important step in successful fume incineration. For these reasons, the distributed raw gas burner is well suited for use in thermal afterburners. These burners are placed directly in the fume stream in order to utilize oxygen in the fume, and required mixing distances can be made short (less time needed for bringing all fume up to the required temperature) since the flame is distributed across the fume inlet to the combustion chamber. Figure 3-3 is a sketch of a system using distributed burners. Drawbacks include: fouling in streams containing tars or particulates; tendency for flame quenching (high CO in combustion products); and inability to burn fuel oil instead of gas in currently available burners.

As is discussed in Chapter 14, subdivision of the cold fume stream and hot combustion products from the supplemental fuel flame is one of the best ways to achieve good mixing in a short distance. The distributed burner divides the flame into many individual jets or lines of flame surrounded by cold fume. Average spacing is 1 foot or less. Thus the scale across which mixing of flame and fume must occur is on the order of inches rather than the full combustion chamber diameter (2 - 10 feet) and mixing time will be reduced. The pollutants have more time at the operating temperature and there is less chance for bypassing.

3.2.1 Description of Available Distributed Burners

There are three basic types of distributed burners utilized in fume incinerators and most are based on burners developed for air heating applications. These are line burners, multijet burners, and grid burners. As will be seen, these are rather arbitrary distinctions and many variations exist. All of these burners are intended as raw natural gas (or propane) burners with air taken from the fume stream, but they can be modified as discussed in Section 3.2.2.2 for use with low oxygen content fumes.

3.2.1.1 Line Burners

Line burners are the most commonly utilized distributed burner. The Maxon Combustifume (Figure 3-4) is widely employed in various manufacturers' afterburners. The Eclipse TAH-0 (Figure 3-5a) and AH-0 (Figure 3-5b) and other specific manufacturers' variations of these burners are also found in fume incinerators in use and on the market today. These are all intended as raw natural gas (or propane) burners with air taken from the fume stream but can be modified for low oxygen content fume as noted below. Gas enters through holes in manifold pipes placed across the duct with spacing on the order of one foot. Air (fume) for combustion enters through holes in mixing plates attached to the manifold pipe and forming a V shaped trough. Lines of flame result.

Figures 3-3 thru 3-7a, b follow

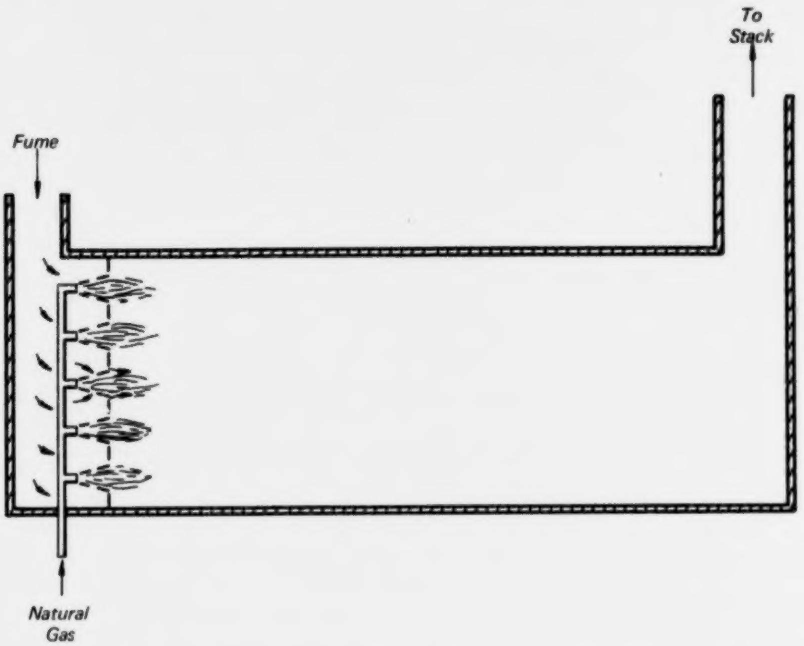

To Stack
Fume
Natural Gas

Figure 3-3. AFTERBURNER WITH DISTRIBUTED BURNER

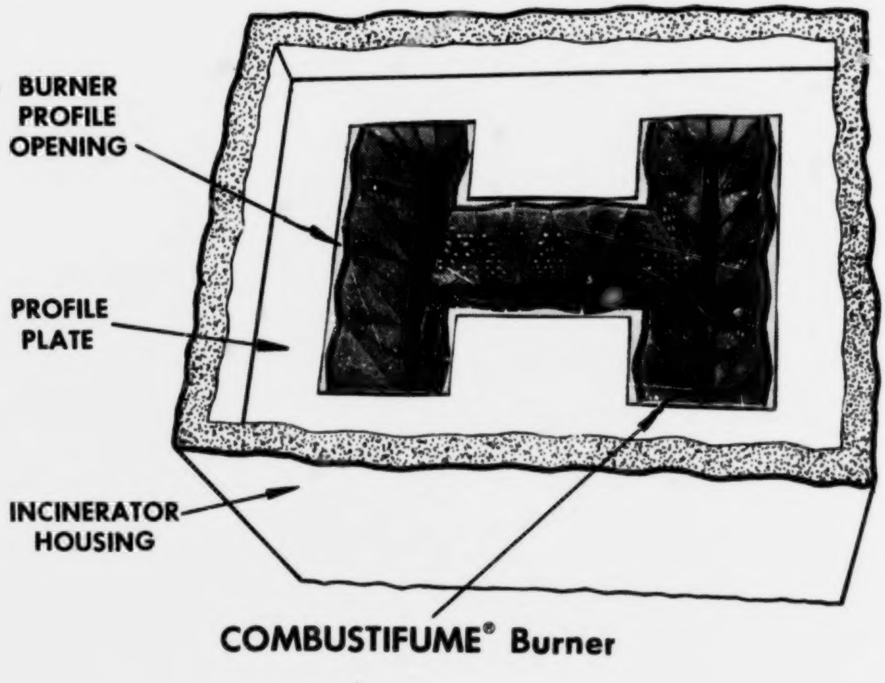

BURNER PROFILE OPENING
PROFILE PLATE
INCINERATOR HOUSING
COMBUSTIFUME® Burner

Figure 3-4. MAXON COMBUSTIFUME BURNER

COURTESY: *Maxon Premix Burner Co.*

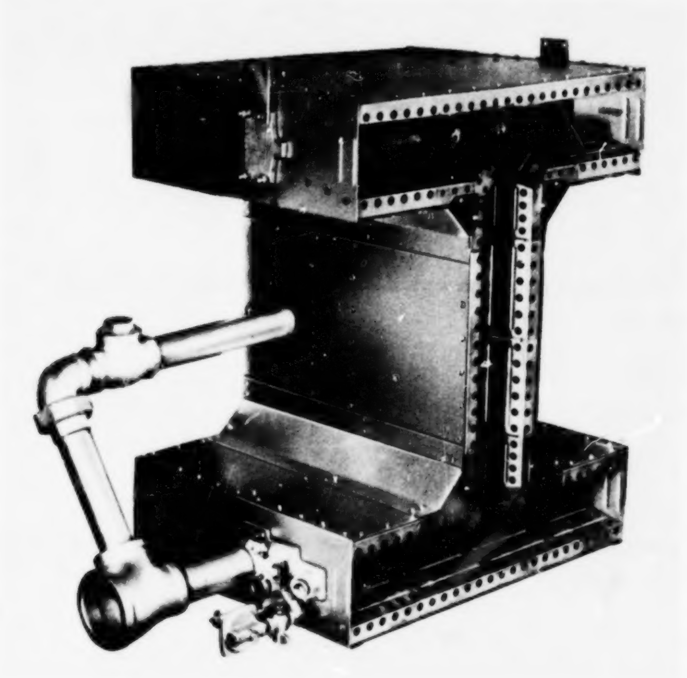


Figure 3-5A. ECLIPSE TAH-O DISTRIBUTED BURNER
Front View - Mixing Plates

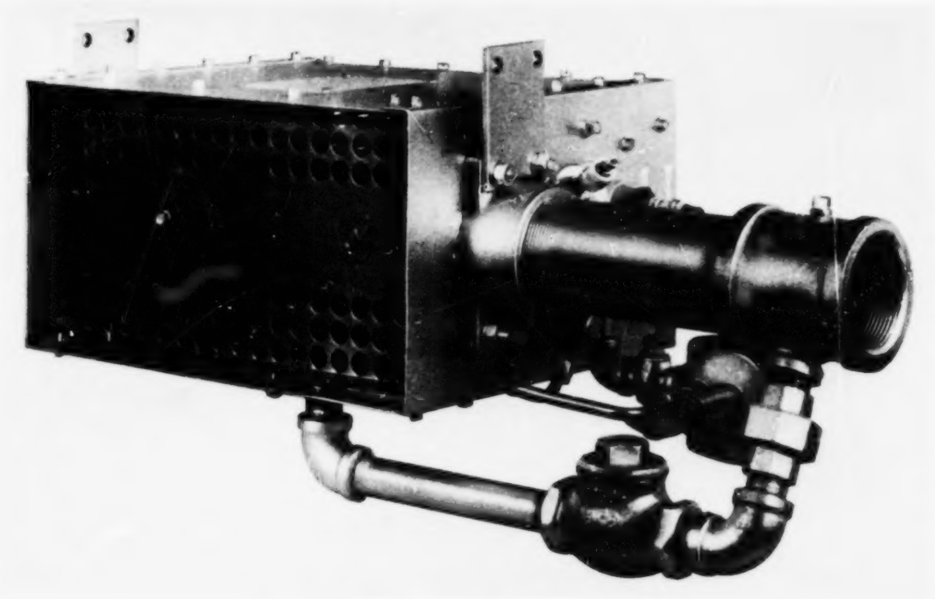


Figure 3-5B. ECLIPSE AH-O DISTRIBUTED BURNER
Back View - Inlet for Fume Used as Combustion Air

30c

COURTESY: *Eclipse Fuel Engineering Co.*

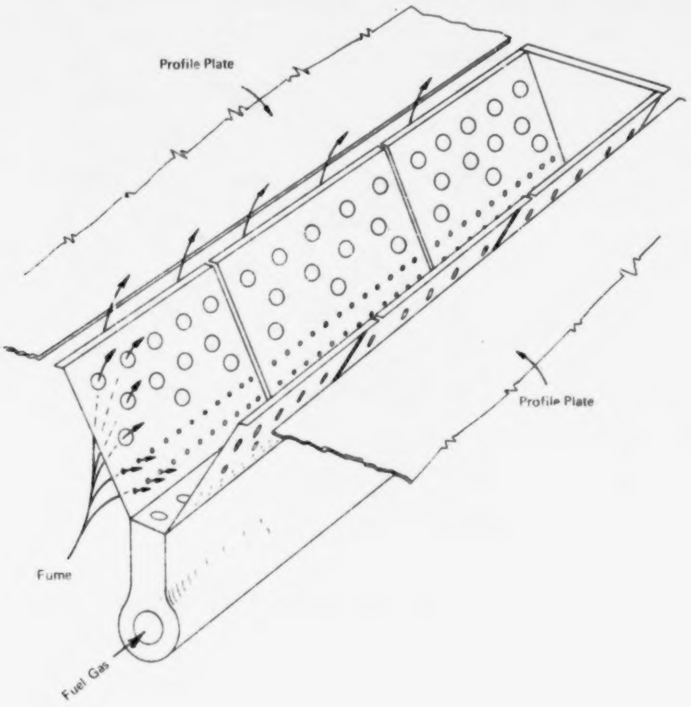
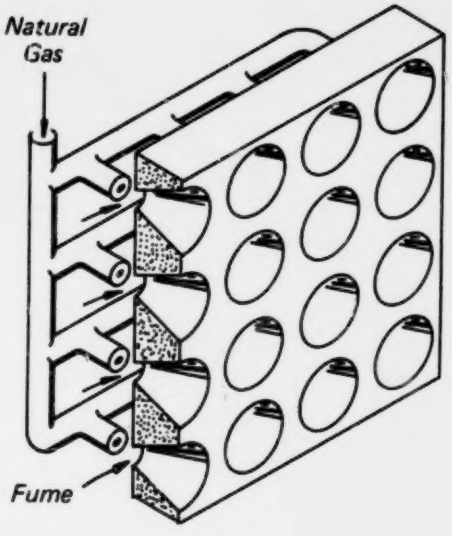

Profile Plate
Profile Plate
Fume
Fuel Gas

Figure 3-6a. COMBUSTIFUME TYPE DISTRIBUTED BURNER WITH PROFILE PLATES

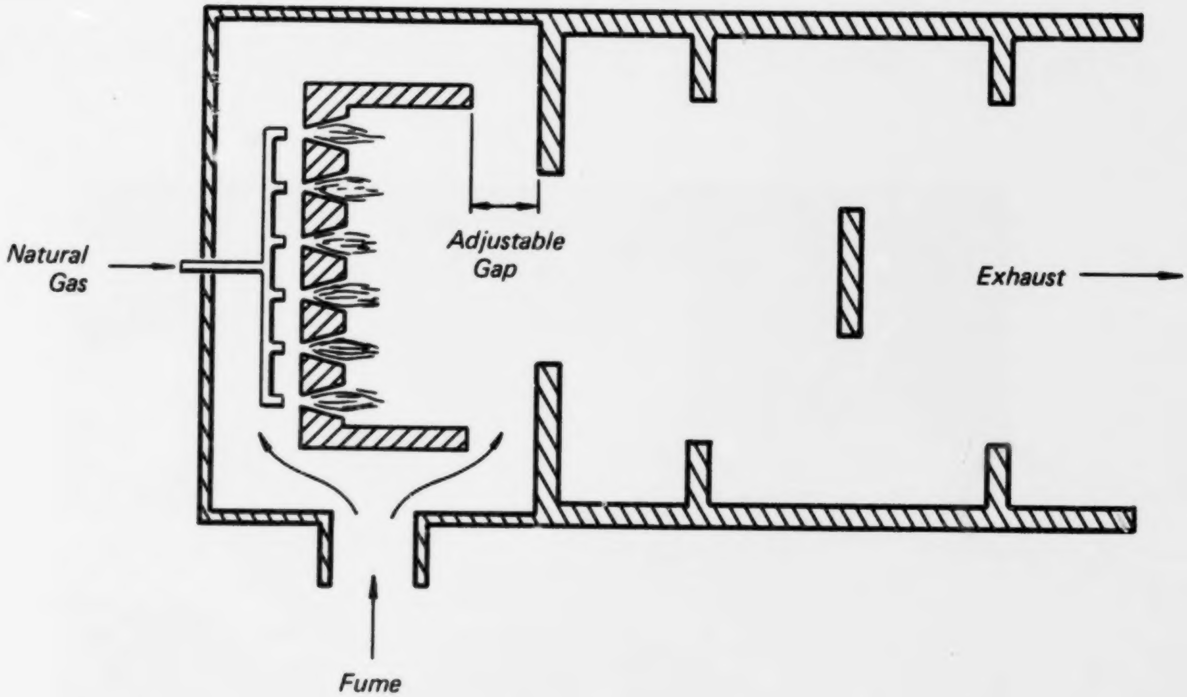
Reproduced from best available copy.

Figure 3-6B. SHORT FLAME FROM COMBUSTIFUME BURNER

COURTESY: *Maxon Premix Burner Co.*


Natural Gas
Fume

A. Hirt Multijet Gas Burner


Natural Gas
Adjustable Gap
Exhaust
Fume

B. Afterburner System Employing Multijet Burner

Figure 3-7. HIRT MULTIJET GAS BURNER

Figure 3-6a shows the Maxon Combustifume type arrangement. Profile plates are used to block most of the area between adjacent burners and between outside burners and the combustion chamber wall. The gap between profile plate and burner determines the pressure drop across the burner and is set to force part of the fume (~50% for reasons discussed in Section 3.1.2.1) through the mixing plate holes to provide oxygen for the flame. The remaining fume bypasses the flame and is mixed with the hot combustion products downstream of the profile plate. The pressure drop across the burner is ~1.5" w.c. which results in fairly intense mixing of natural gas and fume between the mixing plates. This gives a short, stable flame extending only a short distance beyond the V trough, as shown in Figure 3-6b.

The Eclipse type of distributed line burner is similar except that much longer flames are produced and higher heat input per foot of burner is possible. Flames extend up to 5' beyond the burner for the ~1.5 million Btu/hr-ft maximum firing rate. This maximum is ~2X that for the Combustifume type burner and means that burner costs are less for a given heat input requirement (fume flow rate). However, burner spacing will be greater since less burner length is needed, so mixing of fume and flame must occur across somewhat greater distances. Also, the long flame length means that several feet of combustion chamber are lost before all of the heat from the supplemental fuel has been released and can be used to raise the fume to the required incineration temperature (~1400°F). The slow mixing of fume and fuel (which gives the long flame) offers the advantage of somewhat greater turndown (ratio of maximum to minimum firing rate) and spacing between burner and profile plate is less critical. With rapid mixing it is easy to fall outside the combustible limits and get serious flame quenching (Section 3.1.3).

3.2.1.2 Multijet Burners

Multijet burners employ many discrete flames of the type discussed in Section 3.3. However, they are viewed as part of a single burner with one control system and a common manifold for feeding fuel gas. The Hirt multijet burner schematically shown in Figure 3-7a has been widely used in afterburner applications. It is normally installed as indicated in Figure 3-7b. Part of the fume flows around behind the burner, mixes with the fuel jet, and supplies oxygen for combustion. Individual flames issue from each port in the burner. The remaining fume mixes in with the hot combustion products after all of the supplemental fuel has been burned. The gap is adjusted to provide the correct amount of fume flow through the burner; this can be done easily after installation from outside the unit. On the other hand, changing profile plates on line burners is a difficult task that requires shutdown of the afterburner. A drawback to the multijet burner is that bypassed fume (~50% of the total) is brought around the outside of all of the hot combustion products, rather than dispersed throughout as in the line burner. This presents a more difficult mixing problem since the fume must be mixed across several feet of combustion products rather than several inches. Thus, this type of distributed burner is able to use oxygen in the flue gas and avoids problems of flame quenching, but it does not take advantage of the reduced mixing distances possible with distributed burners.

3.2.1.3 Grid Burners

The grid burner presents another method of distributing the flame across the fume duct. The North American Flame Grid Burner is schematically shown in Figure 3-8. Gas enters through holes in the manifold pipes and fume flows through the ~1" X 3" slots in the grid plate. Flames are stabilized by the wake formed behind the grid plate. (See Section 13.2.6). This burner provides extremely fast mixing between fuel and combustion products because mixing distances are very short; manifold pipes are ~6" apart. There is relatively little control of fume and fuel mixing, however, so some quenching will occur if the fume pressure drop across the burner is too high. Also, flame stability is primarily due to the eddies formed behind the grid. If fume pressure drop across the burner is too low, these eddies will be weak and the flame will not be steady. Provision can be made for adjusting the size of the grid openings (a second grid sits over the first and can be moved as a shutter), thereby adjusting the fume pressure drop. However, this grid burner is most suitable for applications where fume flow rate is relatively constant.

Figure 3-8 follows

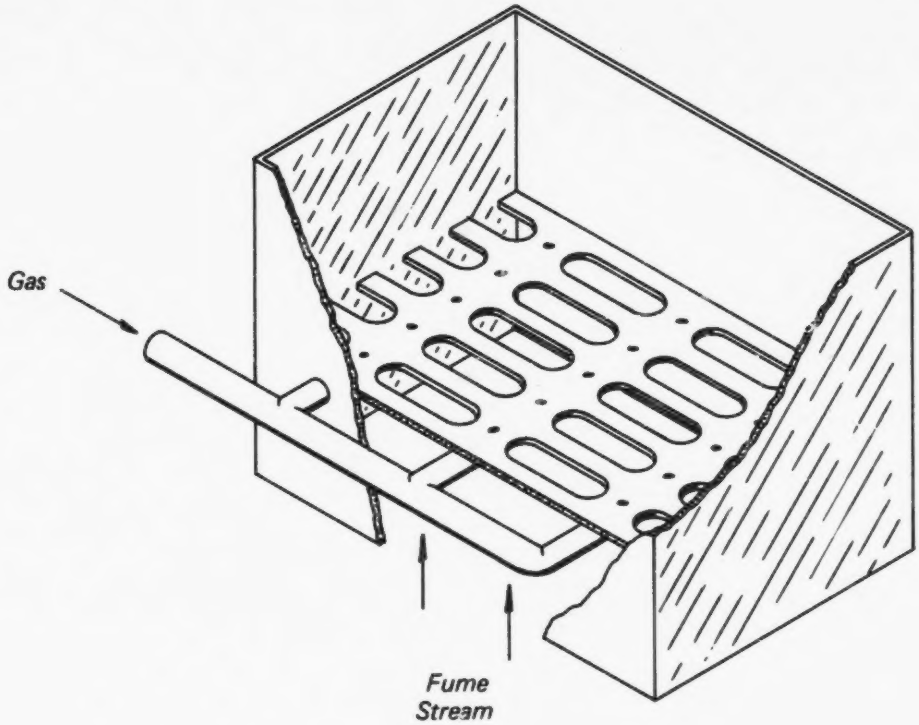

Gas
Fume Stream

Figure 3-8. NORTH AMERICAN FLAME GRID BURNER

3.2.2 Possible Problems with Distributed Burners

Possible problems with flame quenching due to rapid mixing with excess fume have already been discussed. Other problems are primarily related to the fact that distributed burners are placed directly in the fume stream.

3.2.2.1 Burner Fouling

If liquid or solid particulates are entrained by the fume, buildup on the burner can occur, resulting in plugging various burner parts. Mixing plates, burner tiles, and grid plates are generally quite hot due to heat transferred from the flame so combustible material will usually burn off. However, buildup in cooler areas can occur without burnoff (or noncombustible material may be involved) and this can lead to plugging. More serious is the possibility that combustible deposits will accumulate and then ignite, releasing enough heat to damage the afterburner. Distributed burners should not be used in fouling streams unless frequent cleaning is possible.

3.2.2.2 Low Oxygen Content

Another limitation on some distributed burners is that fume oxygen content should be above 16%. Lower values are said to be insufficient to guarantee complete combustion of the supplemental fuel. Oxygen content was well above this for most processes encountered in our afterburner user survey. Anytime air dilution is used to keep below 25% of the LEL, oxygen content is expected to be well above 16%, even in the case where afterburner exhaust is recycled to the process for direct heat recovery (Section 6.2.4). Batch operated processes such as wire burnoff, coffee roasting, and refuse incineration have a tendency to produce low oxygen content fumes during part of the cycle. Certain processes involve dilution by N_2 or CO_2 and fume oxygen content is near zero.

Low oxygen content can be partially overcome in any of the distributed burners by partially premixing air with the natural gas fuel. The burners are not designed for completely premixed operation since flammable mixtures would be contained in the manifold but ~25% of the air needed for combustion can be added with minor modifications. (One cubic foot of methane requires 9 ft^3 of air for complete combustion so ~2.25 ft^3 of air can be premixed. This gives a ~30% methane mixture). In addition, the multijet burner and the Eclipse type burner can be readily modified to utilize outside air for combustion. All fume is then forced to bypass the flame.

It would also be possible to add air directly to the fume stream to provide sufficient oxygen for combustion. However, as is shown in Table 3-3, this is not very efficient since much more air is needed to achieve the 16% minimum oxygen content in the fume than is needed through partial premixing or fuel combustion with outside air.

Table 3-3

Fume Oxygen Content	Ft^3 Of Outside Air Needed Per Ft^3 Of Fume To Achieve 16% O_2	Ft^3 Of Outside Air Per Ft^3 Of Fume If Fuel Burned With Outside Air And All Fume Bypassed	
		25% Excess Air (OK For Discrete Burner But Low For Distributed)	100% Excess Air
16%	0	.55	.87
10%	1.2	.55	.87
0%	3.2	.55	.87

Thus, large amounts of outside air would be needed for direct addition to the fume and supplemental fuel must be burned to heat all of this air. In general, discrete burners will be able to achieve complete combustion with less excess air and therefore will show better fuel economy when outside air must be used.

3.2.2.3 Excessive Temperatures - Limit on Preheat

Distributed burners are subject to high rates of heat transfer from the flame so mixing plates and grid plates tend to reach high temperatures. The incoming fume provides some cooling of the plates but metal temperatures of 1800°F or higher have been observed. This temperature will increase with increased preheat temperature, increased combustion chamber temperature and increased pressure drop across the burner. High temperature alloys such as Hastelloy X are generally utilized in burners available today. Many failures have occurred with aluminized steel and stainless steels such as type 304. Even Hastelloy X can be damaged if combustion chamber temperatures exceed 1500°F and sufficient cooling is not provided by the fume. Refractory construction, such as in the multijet burner, imposes no limits at temperatures of interest.

Excessive preheat of fume can also cause decomposition of natural gas in the distributor manifold, producing coke which plugs the gas orifices. This becomes a problem at preheat temperatures greater than 1000°F. Such high temperatures can occur if fume is raised to a temperature where oxidation of the hydrocarbons in the fume occurs at a rapid rate and much of the fume's heat content is released upstream of the burner.

3.2.3 Afterburner Configuration - Mixing of Bypassed Fume

Once supplemental fuel has been burned to generate heat for dilute fume incineration, the bypassed fume must be mixed with the hot combustion products and then held at the required temperature for sufficient time. Both mixing and retention (Section 3.1.1) occur within the combustion chamber of the afterburner and are difficult to distinguish from one another. Here we discuss combustion chamber configuration only as it affects the mixing step since this is dependent on whether distributed or discrete burners are used. Design features pertaining to uniform retention of pollutants at the incineration temperature and most mechanical considerations are discussed in Section 3.4 since these are the same regardless of burner type.

Afterburners should be designed to give rapid and complete mixing of fume and hot combustion products once the fuel has been combusted. Flame quenching should be minimized and a balance must be struck between expenditure of pressure drop to speed mixing and the added purchase cost for additional total residence time in the combustion chamber. Chapter 11 contains cost information to guide this decision. With afterburners designed to give a certain total residence time (since we cannot predict mixing processes well enough for use as an absolute design criterion), the afterburner configuration plays a major role in deciding how efficient the unit will be at pollutant destruction.

3.2.3.1 Location of Distributed Burner

The distributed burner is placed directly in the duct carrying the fume stream and fires into one end of the afterburner combustion chamber. It is important that fume flows uniformly through the entire cross section of the burner in order to achieve a nearly uniform temperature profile a short distance from the burner. The typical 1 - 1.5" w.c. pressure drop across distributed burners will smooth out modest variations in fume velocity across the inlet duct. However, wherever possible a straight section of inlet duct extending at least 3 - 4 duct diameters should be placed between any 90° bend and the burner. This is especially important where flow variations can cause the burner pressure drop to fall below 1/2" w.c. Figure 3-21 indicates the recirculation or stall region which is set up downstream of a sharp bend. If the burner is placed within this region, gross maldistribution may result. (The pressure drop over the burner has a strong effect on the location of such a recirculating region so the sketch should be taken only as an indication of potential problems.)

3.2.3.2 Cross Sectional Area

Combustion chamber cross sectional area should be set to insure that flows will be strongly turbulent so that mixing will occur rapidly on the molecular level and pollutant molecules will be heated and oxidize. Turbulent flow requires

$$\frac{(\mathrm{SCFM})(\mathrm{D})}{\mathrm{A}} \gtrsim 120 \qquad (3\text{-}2)$$

or for a circular cross section,

$$\frac{\text{SCFM}}{\text{D}} \gtrsim 94 \tag{3-3}$$

where SCFM = Total volumetric flow rate at 60°F and atmospheric pressure (Take fume flow where outside air not used for combustion), Ft^3/min

D = Diameter (or equivalent), Ft

A = Interior cross sectional area, Ft^2

This is an easy criteria to meet and is equivalent to a minimum velocity of 3.5 ft/sec for a 2' diameter combustion chamber. Minimum velocities several times this high (10 - 20 ft/sec) should be utilized to help intensify mixing on the molecular level. For a given total afterburner residence time the best mixing will be obtained for the smallest cross section and therefore the longest combustion chamber. High length to diameter ratio means high velocities and high pressure drops but this results in better mixing and better pollutant destruction. Typically, afterburner design velocities are ~30 ft/second but L/D ratios are only 2 - 6. Cross channel mixing is very slow in straight duct flow and mixing lengths equivalent to 40 duct diameters are required for typical flow rates. Therefore, some additional means is required to accelerate cross channel mixing in order to raise all fume to the incineration temperature in a reasonably sized afterburner.

Chapter 14 discusses the various means of improving cross stream mixing of fume and combustion products. Axial jets can be-used and give mixing distances proportional to the width of the streams to be mixed. With distributed burners the bypassed fume and flame streams are normally greatly subdivided and mixing occurs in a distance shorter than the length of the afterburner. Reasonably complete mixing (see Section 14.2) occurs in a distance about 10 - 20 times the width of the streams. For a one foot spacing this takes 10 - 20 feet.

3.2.3.3 Baffles

It is common practice to employ line and grid burners without any downstream baffles or other mixing promoters. If the flame is sufficiently subdivided and flame and bypassed fume issue from the burner with a large velocity difference, jet mixing will be rapid and sufficient pollutant destruction will normally occur. However, problems can occur if flow through the burner is non-uniform (Section 3.2.3.1) so that mixing must occur across the complete channel width. The burner should be located so that this does not occur.

The layer of fluid flowing along the walls of the duct will have a lower velocity and experience little mixing with the main body of fluid. Therefore, it is likely that a certain amount of fume will flow around the outside of the burner, reach the combustion chamber walls, and pass through the combustion chamber without

Figures 3-9, 3-10 follow

S-14121
67784

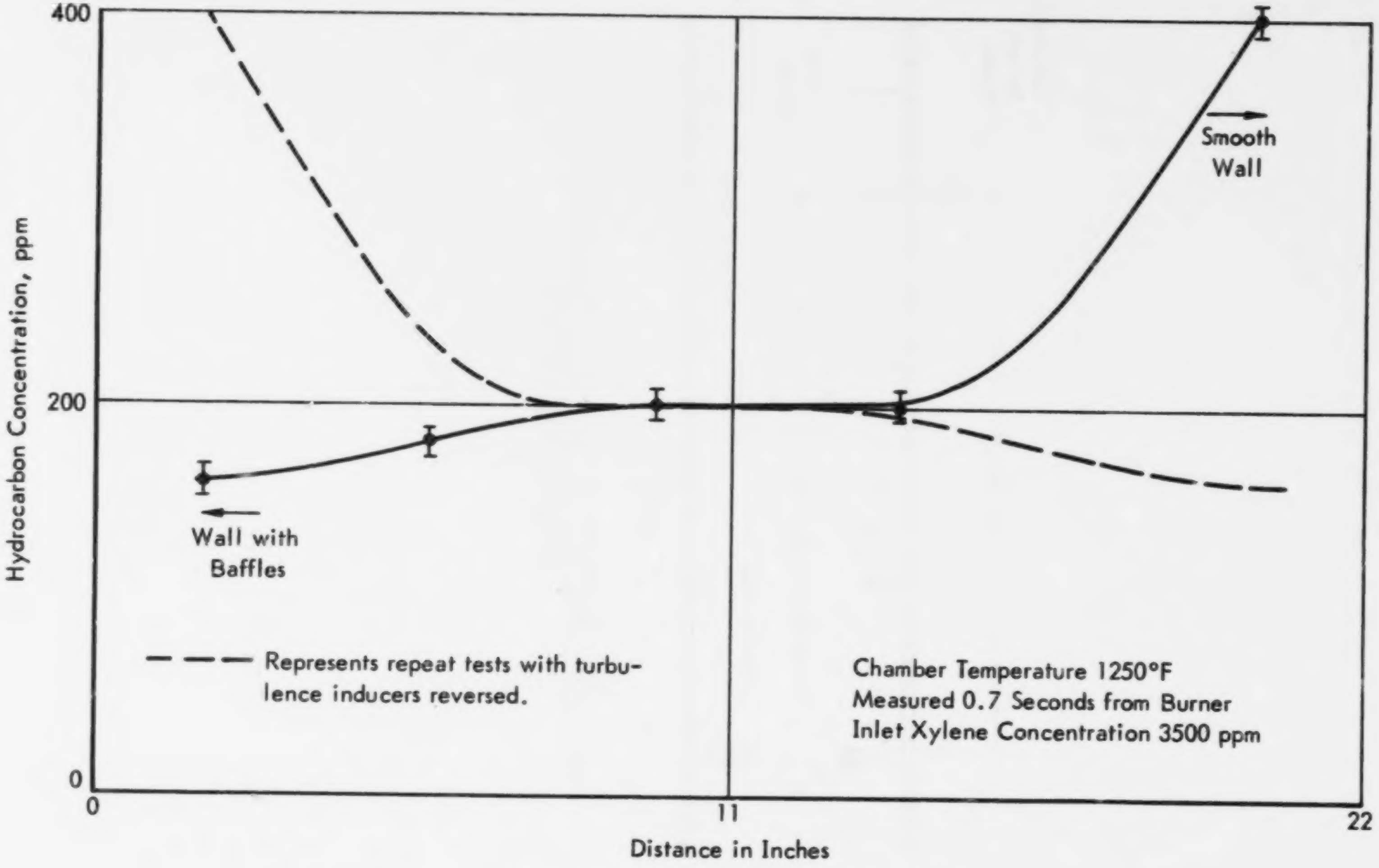


Figure 3-9. EFFECT OF WALL BAFFLES

Courtesy: Despatch Oven Company

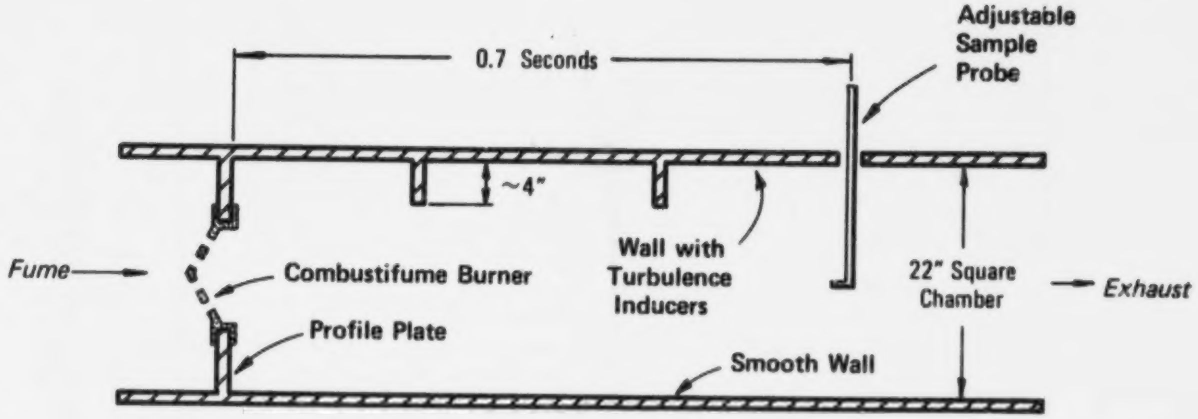

Adjustable
Sample
Probe
0.7 Seconds
~4"
Wall with
Turbulence
Inducers
Fume
Combustifume Burner
22" Square
Chamber
Exhaust
Profile Plate
Smooth Wall

Figure 3-10. DESPATCH OVEN CO. EXPERIMENT TO STUDY TURBULENCE INDUCERS

attaining a sufficiently high temperature for pollutant destruction. The amount of such wall flow depends on flow velocity, burner configuration, chamber configuration, etc., but it is usually small enough so that overall efficiency can be attained with a small increase in combustion chamber temperature.

Ring baffles or "turbulators" placed along the wall serve to divert this wall flow and mix it with the main stream. Figure 3-9 is based on data taken by Despatch Oven Co. showing the effect of such wall baffles on hydrocarbon concentration .7 seconds downstream of a Combustifume burner with the nominal combustion chamber temperature set at 1250°F. The experimental set-up is schematically indicated in Figure 3-10. Approximately 4" wide baffles were placed along one wall of the 22" square chamber. Concentrations were measured at various points across the combustion chamber and considerable variation was observed between concentrations at the "rough" and smooth walls.

Checkerwork baffles (bricks layed so that a gridwork is formed) are also utilized with some line burners to generate additional turbulence and obtain a uniform temperature profile more quickly. As noted in Section 14.8 such grid baffles are quite effective at achieving mixing over the scale of the grid, ~6", but are nearly worthless if larger scale maldistribution exists upstream of the baffle. For this reason, grid baffles with <50% free area are effective in improving mixing when distributed burners are utilized and will shorten the required total residence time. All baffles should be located so that flame impingement does not occur. As is noted in Chapter 7 such flame impingement can set up severe thermal stresses and lead to failure even if the materials' temperature limit has not been exceeded. In addition, partial flame quenching can occur.

As shown in Figure 3-7b the multijet burner normally employs baffles to improve mixing between bypassed fume and hot combustion products since the streams are not subdivided. The mixing problems are then similar to those in afterburners employing discrete burners, Section 3.3. Good cross stream mixing is provided by forcing the bypassed fume to enter at right angles to the combustion gases and having an expansion immediately after the first baffle. The target or disc baffle within the chamber forces another change in flow direction and provides good mixing over large radial distances. To be effective these baffles must block a large fraction of the flow area (~1/2) and this will add at least 1 - 2" w.c. to the system pressure drop. This degree of baffling is often not utilized and poor performance results unless excessively high temperatures are employed. The data shown in Table 3-4 show the importance of proper use of baffles. Inlet concentration variations and possible burner adjustments make direct comparison difficult, but operation with the baffle is obviously better.

Table 3-4. EFFECTIVENESS OF BAFFLES

Multijet Burner - 1400° Combustion Chamber Temperature
.3 Seconds Residence Time

	Without Baffle	With Baffle
Inlet		
Hydrocarbons	900 - 1200 ppm	1300 - 1400 ppm
Carbon Monoxide	10 ppm	30 ppm
Outlet		
Hydrocarbons	250 - 700 ppm	100 - 110 ppm
Carbon Monoxide	20 - 100 ppm	~10 ppm
Maximum-Minimum Temperature Found in Traverse Across Combustion Chamber	120° - 240°F	~30°F

3.3 Thermal Afterburner Design Features - Systems Utilizing Discrete Burners

In Section 3.2 the distributed burners' subdivision of fume and hot combustion products was listed as a major advantage since all fume can be raised to the incineration temperature in a reasonably short distance from the burner. However, there were also disadvantages (fouling, limited turndown, can't burn oil, difficult to use outside air, etc.) and these can usually be overcome in systems utilizing discrete burners. In addition, discrete burner costs are usually lower per million Btu's of heat added, but as noted in Chapter 11, burner costs are a very small factor in total afterburner cost. Figure 3-11 is a sketch of a typical afterburner utilizing a discrete burner.

Our distinction between discrete and distributed burners is somewhat arbitrary. The discrete burner is viewed as a point source of flame with a separately controlled fuel supply. There may be four or more such burners in a large afterburner but they are distinct, not like the multijet burner (Section 3.2.1.2). They may be raw gas burners which sit in the fume stream and obtain oxygen for combustion directly from the fume. Ring burners and one burner employing mixing plates similar to the line burners of Section 3.2.1.1 are classified as discrete burners. The commonly used nozzle mix or "torch" burners with their long flames are discrete burners, but intense air/fuel mixing may also be employed to give a very short flame. Discrete burners may utilize a completely premixed stream of natural gas and air as well as raw gas or oil fuel.

The major distinction is that the burners are discrete sources of heat and additional mixing measures must be employed to prevent fume passing through the combustion chamber without being raised to a sufficiently high temperature for pollutant destruction. Single jet mixing and turbulent cross stream mixing are not effective in afterburners with typical length to diameter ratios of 2 - 6. Many existing afterburner designs rely completely on mixing provided by energy in the flame jet. This has led to poor performance and is especially a problem when firing rate is low because of high fume preheat or high solvent content. High pressure drops (>4" w.c.) and combustion chamber baffles must be used to obtain sufficient mixing and adequate performance with total residence time of .5 seconds or less.

3.3.1 Discrete Burners Utilized in Available Afterburner Systems

All of the major burner manufacturers make gas and oil burners which can be and are used in fume incinerators currently on the market. There are many variations among these burners. These variations affect flame shape, turndown ratio, and completeness of combustion. Chapter 13 and burner suppliers should be consulted for guidance in choosing between different burners. In general, poor afterburner performance is not due to the burner itself, but rather to how it is utilized in the complete design. The burner is viewed as a source of heat and if properly adjusted it will provide relatively complete combustion, at least near the design firing rate.

3.3.1.1 Raw Gas Burners - Nozzle Mix with Low Pressure Primary Air Supply

Since natural gas is clean burning and readily available at a low price in most areas of the country it is most often used for supplemental fuel in an afterburner. The most common arrangement employs a jet of gas with primary combustion air brought in around this gas jet to initiate combustion. This primary air in the nozzle mix type of burner is usually fresh air and set to supply no more than the minimum oxygen needed to complete combustion. Additional oxygen is taken from the fume stream. In order to reduce the heat load by as much as 30%, part of the fume stream (if it will not foul or erode burner parts) could be supplied as primary air for combustion instead of adding outside air. However, only one or two installations were encountered in which this was done. Combustion air is supplied at ~0.5" - 4" w.c. pressure and must mix with the gas stream to form a combustible mixture. This mixing is very slow for low supply pressures so the resulting flame is typically several feet long and not well defined. If excess fume is mixed in too rapidly, serious flame quenching can occur, but simple entrainment of fume into the flame jet should give no such problem. Turndown on fuel is limited to ~5:1 with no adjustment of the burner. No automatic control of air supply is employed. (See Section 3.3.1.3.)

3.3.1.2 Raw Gas Burners - All Oxygen From the Fume

The two raw gas burners discussed here operate much like distributed burners except that they occupy a small fraction of the combustion chamber cross section and are classified as discrete. Fuel savings associated with use of oxygen in the fume result, but the problems associated with placing distributed burners in the fume stream (Section 3.2.2) also are possible. In many cases a simple gas ring has been placed in a fume stream, especially in an exit stack in order to eliminate a visible plume. Problems inherent in such an installation have been discussed in Section 3.1.4.5. A simple raw gas ring burner will add much CO and unburned fuel to the fume unless flame stabilization is provided and mixing with fume is controlled. Partial flame quenching is less of a problem in a 1400°F afterburner with uniform treatment of fume since CO and aldehydes from the flame will be oxidized along with pollutant. Figure 3-12 shows one afterburner design employing a ring burner. The stabilization bar provides some control over the rate of fume mixing to avoid quenching. In this afterburner design, a mixing baffle downstream of the stabilizer provides fairly uniform treatment of the entire fume stream.

UOP uses a raw gas burner with a conical mixing plate in their thermal fume incinerators. Figure 3-13 is a sketch of this burner. Outside air is normally not used and fume is split between combustion air and bypass according to the pressure drop across the burner. Operation is like the Combustifume type line burner (Section 3.2.1.1) but a single flame jet results and extends several feet into the combustion chamber. As in the line burners, high temperature alloys must be used for construction of the mixing plate or cone.

3.3.1.3 Raw Gas Burners - High Turndown Capability

Where high turndown of the supplemental fuel firing rate is desired, a special wide range burner should be used. These burners control both air and fuel supply and can achieve a 40:1 or greater variation in firing rate without much loss of stability. Combustion air must be supplied at ~1 psig. Such a large turndown is desirable in cases where the fume flow rate or composition is highly variable (This makes successful afterburner operation difficult, however.) or

Figures 3-11 thru 3-15 follow

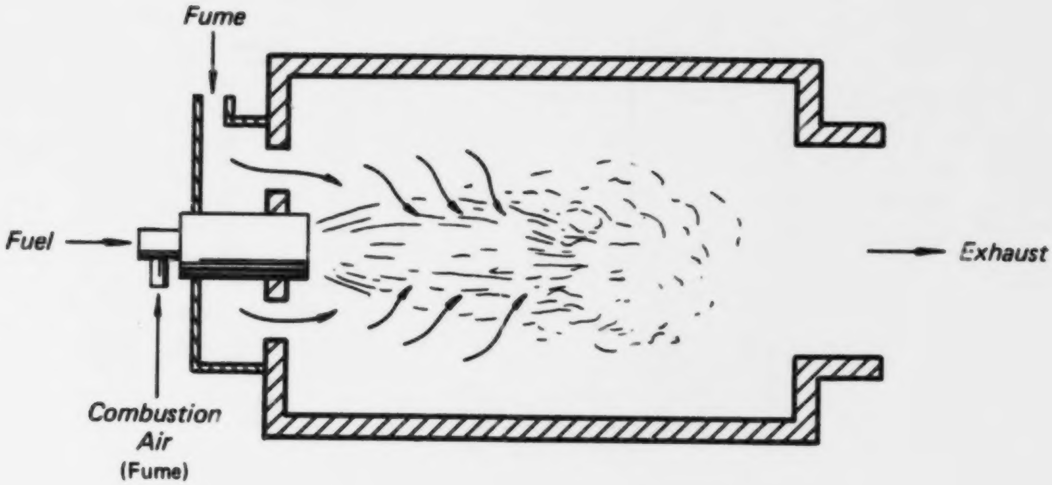

Fume
Fuel
Exhaust
Combustion
Air
(Fume)

Figure 3-11. AFTERBURNER USING A DISCRETE BURNER

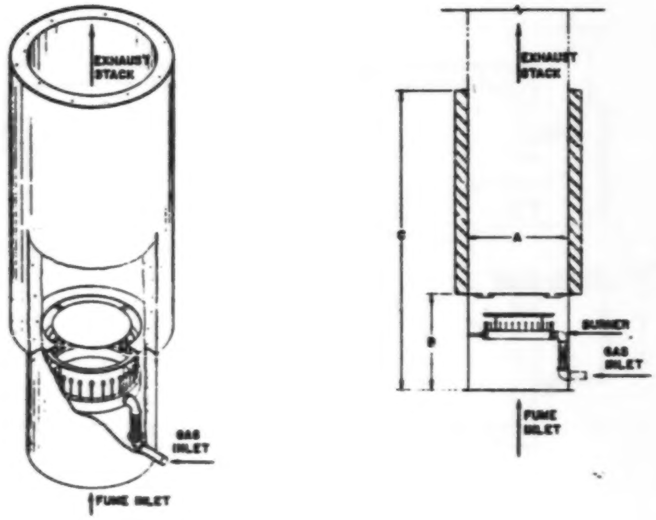

EXHAUST STACK
GAS INLET
FUME INLET
EXHAUST STACK
A
C
B
BURNER
GAS INLET
FUME INLET

Figure 3-12. RING BURNER USED IN STACK AFTERBURNER

COURTESY: *Surface Combustion Division – Midland Ross Corp.*
& *The Heating and Ventilating Engineer Journal* 12-9)

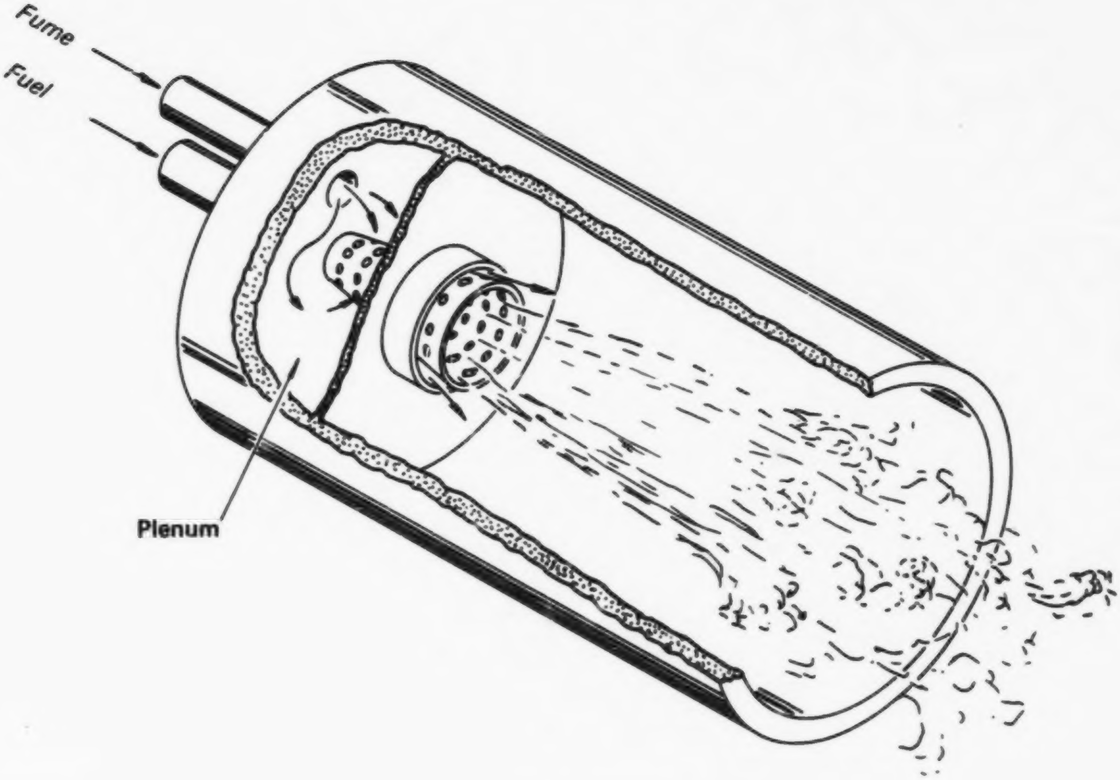

Fume
Fuel
Plenum

Figure 3-13. UOP RAW GAS BURNER

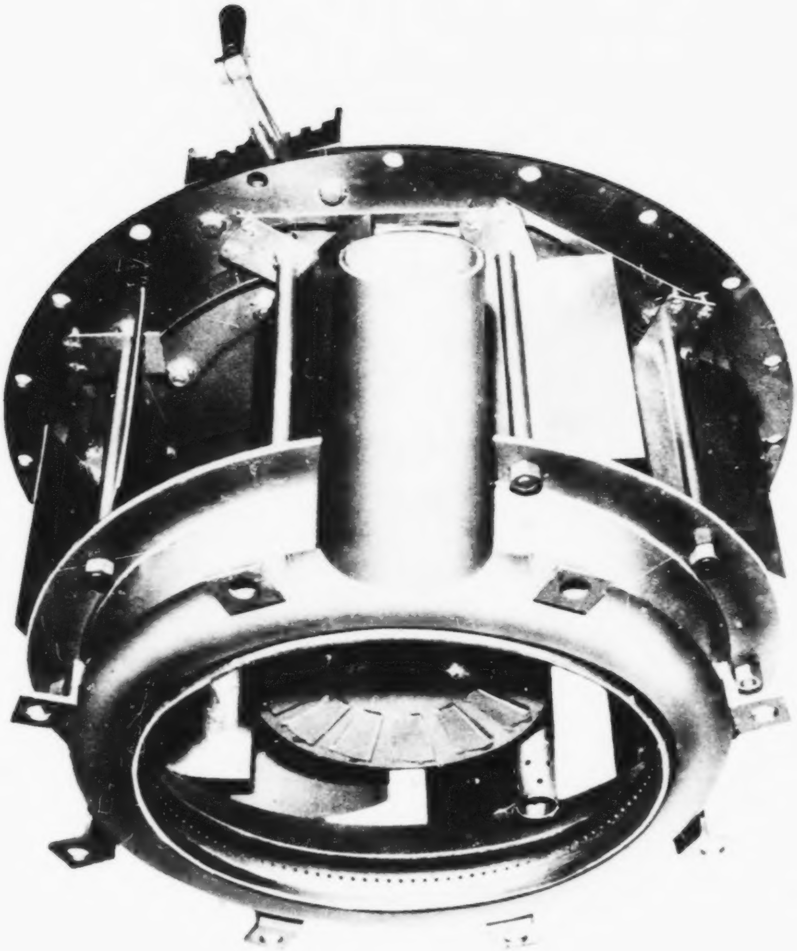


Figure 3-14. COMBINATION OIL AND GAS BURNER EMPLOYING SWIRL

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COURTESY: *Combustion Equipment Associates*

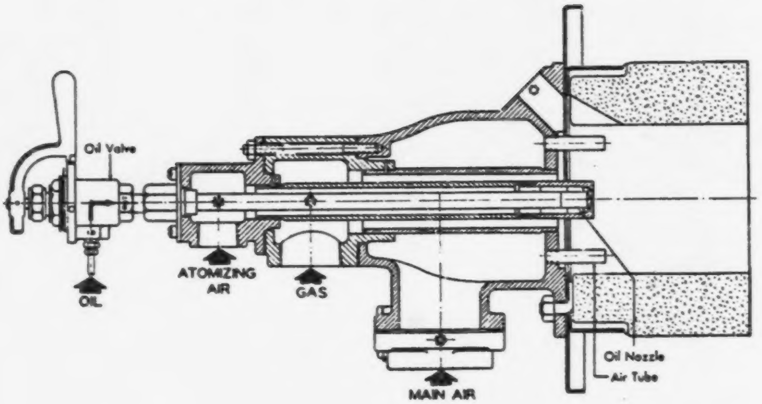
Oil Valve
ATOMIZING AIR
GAS
OIL
MAIN AIR
Oil Nozzle
Air Tube

Figure 3-15. NORTH AMERICAN DUAL FUEL BURNER

COURTESY: *North American Manufacturing Co.*

highly efficient heat recovery is employed and a high firing rate is required only for start up. High turndown can also be achieved by using multiple burners so that one or more can be shut off when not required. Fume can be used for combustion air, but normally fresh air is used to avoid fouling of the blower and burner.

3.3.1.4 Premix Gas Burners

The kitchen gas range is a common example of a premix flame issuing from a ring burner. Gas and air are mixed up stream of the burner and a relatively short, stable flame results. (See Section 13.2.3.1.) Where premix burners are used in afterburners they have usually been a single nozzle with the air/gas mixture produced in a premixing blower. Outside air is used since burner ports are small and subject to fouling. High turndown is possible (10:1 - 15:1 or more) since both air and fuel are easily varied. Premix burners are not normally utilized for heat duties above one half million Btu's/hour but rapid heat release and short flames are attainable.

3.3.1.5 Oil Burners and Dual Fuel Burners

In certain areas natural gas is difficult to obtain and/or it is very expensive. In such cases oil fired afterburners are attractive or one may want the capability to substitute oil fuel during periods where the natural gas supply is interrupted. Gas shortages are a relatively recent phenomenon so many afterburner suppliers do not regularly supply an oil fired option. However, most designs employing discrete burners can be adapted to oil fuel. Oil firing increases flame luminosity and therefore will transfer more heat by radiation. (However, the fume, unless it contains particulates, will not absorb much heat by radiation. The chamber walls and any other solid objects which "see" the flame can be expected to exceed the average chamber temperature unless cooled from behind.) Also for similar burner mixing patterns, oil flames can be expected to be longer than gas flames by 50% or more since the oil must first vaporize. These points are discussed in Sections 13.2.4 and 13.3.2.1. Maintenance requirements will be higher with oil burners than gas burners.

Figure 3-14 shows a typical combination oil and gas burner. Raw gas enters through the ring at the exit end of the burner. An oil gun is inserted through the hole in the center of the burner. Combustion air or fume enters through the louvers along the outside of the burner. The louvers are set to give some spin to the air giving more intense mixing and a shorter flame. Spin is discussed in the next section. The burner is mounted on the end of the combustion chamber and surrounded by a "windbox" which directs fume or air through the louvers. The slotted plate surrounding the oil gun serves to stabilize the flame by recirculation and spin (Section 13.2.6). Figure 3-15 shows the commonly used North American Dual Fuel Burner which does not use spin on the combustion air.

Oil fuels must be sprayed into the chamber as small droplets so the vaporization is fast and combustion complete. Typically, afterburners employ #2 fuel oil, a relatively low viscosity home heating oil. It is "pressure atomized" through spray nozzles with oil supplied at 25 - 100 psig. Alternatively, oil can be supplied at low pressure and air jets used for atomization. In some cases heavy residual (#6) oil may be utilized and this requires heating to lower its viscosity and use of high pressure steam or air jets to break up the oil into small droplets. (Section 13.2.4 discusses oil atomization in more detail.) Residual oil is cheaper but it normally is high in sulfur and nitrogen which will end up in the afterburner flue gas so cleaner fuels are preferable.

Oil burners including their atomizers have a limited turndown ratio of 3:1 - 4:1. Pressure drop on the fume or air side ranges from 5" - 10" w.c. for the type shown in Figure 3-14. Lower pressure drop operation (~2" - 4" w.c.) is possible with burners similar to low pressure raw gas burners but flame lengths are much longer and the supplemental fuel heat of combustion is not available for fume heating until far down the combustion chamber. If oil burners are used, it is crucial that oil guns are blown out and pulled from the burner immediately after the oil flow is stopped. Otherwise carbonization will occur and burner plugging and poor atomization will be a problem.

3.3.1.6 Use of Swirl To Obtain Rapid Heat Release

As noted in Section 3.1.3 little benefit is to be derived from "flame contact" whereas partial flame quenching leading to high CO, aldehydes, and unburned fuel in the flue gas is quite likely. Therefore, supplemental fuel should be combusted as rapidly as possible to provide maximum residence time for mixing with bypassed fume and retention of pollutants at the incineration temperature. A precombustion chamber with careful control of combustion air (fume) would insure complete combustion prior to entering the afterburner combustion chamber.

Intense swirl or vortex burners can be used to speed up air/fuel mixing and achieve supplemental fuel combustion with a very short flame. Figure 3-16 is a sketch of a vortex type burner. Combustion air (which can be part of the fume stream if it contains sufficient oxygen and isn't fouling) is supplied at high pressure (~16 - 20" w.c.) and much of this pressure is lost in passing through the swirl vanes (very tight louvers such as in Figure 3-14). Air then passes around the fuel jet with a high rate of spin and expands into the precombustion chamber. This spin and expansion leads to the circulation patterns indicated in Figure 3-16. Mixing and combustion are rapid and very little flame extends beyond the end of the precombustion chamber which is considered part of the burner. As is discussed in Sections 13.3.2.3 and 14.3.2.5 flame shape and mixing intensity depend on the ratio of axial velocity to swirl velocity at the inlet to the burner.

Such a swirl burner can be used to provide a source of hot combustion gases without danger of flame quenching provided that the air/fuel ratio stays within combustible limits (Section 12.1.2). Mixing is nearly complete so one must be extremely careful that too much excess air (or fume) is not put through the burner and that oxygen content remains high enough. Typically, such burners are run with close to the stoichiometric amount of combustion air so high temperature (~3000°F) combustion gases result. If part of the fume stream is used for combustion air it must be passed through a blower to provide the required high pressure (~16" - 20" w.c.). Fouling and erosion are potential problems. Because of the sensitivity of mixing patterns to flow rate, turndown is limited to ~3:1.

Figures 3-16, 3-17 follow

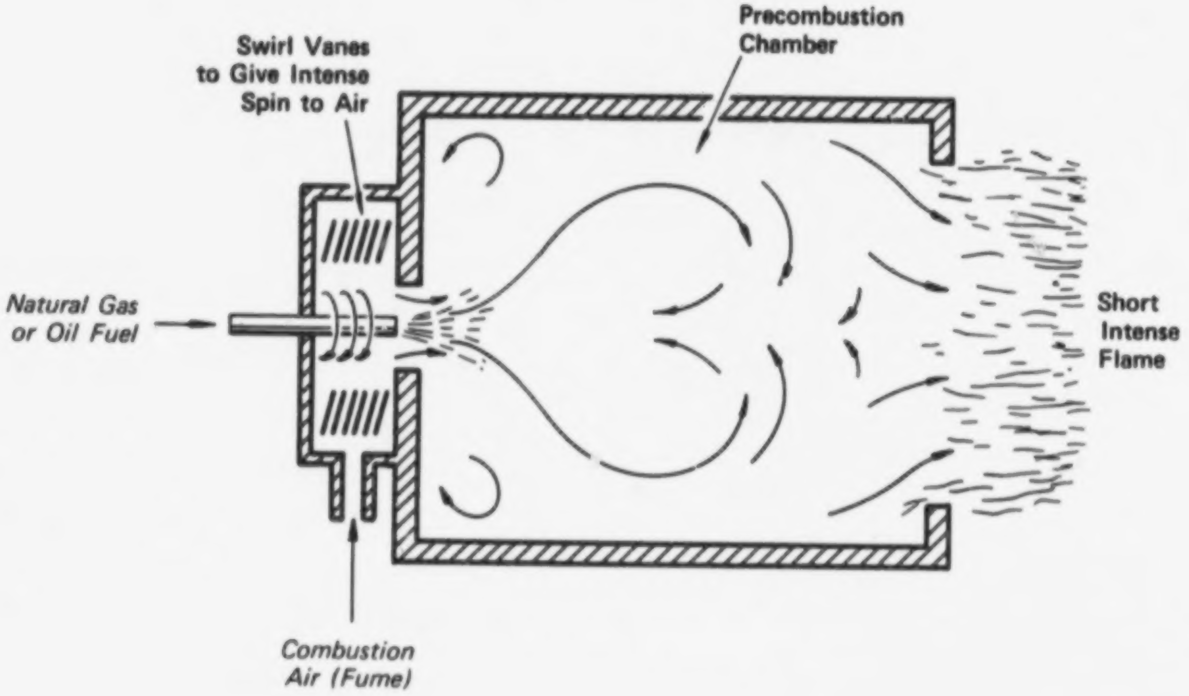
Precombustion Chamber
Swirl Vanes to Give Intense Spin to Air
Natural Gas or Oil Fuel
Short Intense Flame
Combustion Air (Fume)

Figure 3-16. VORTEX TYPE BURNER-RAPID AIR FUEL MIXING

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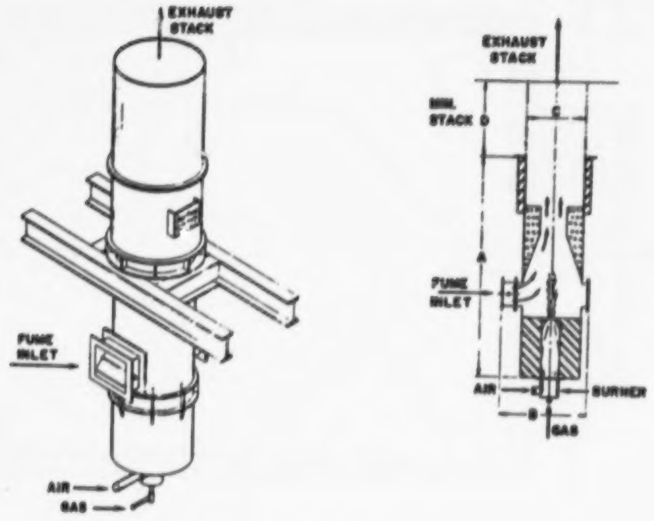

EXHAUST STACK
FUME INLET
AIR
GAS
EXHAUST STACK
MIN. STACK D
C
A
FUME INLET
AIR
E
BURNER
B
GAS

Figure 3-17. USE OF FLAME JET TO INDUCE FUME FLOW

COURTESY: *Surface Combustion Division – Midland Ross Corp.*
& *The Heating and Ventilating Engineer Journal* (29)

3.3.2 Afterburner Configuration - Mixing of Bypassed Fume

Afterburners should be designed to give rapid and complete mixing of fume and hot combustion products once the fuel has been combusted. Flame quenching should be minimized and a balance must be struck between expenditure of pressure drop to speed mixing and the added purchase cost for additional total residence time in the combustion chamber. Chapter 11 contains cost information to guide this decision. With afterburners designed to give a certain total residence time (since we cannot predict mixing processes well enough for use as an absolute design criteria), the afterburner configuration plays a major role in deciding how efficient the unit will be at pollutant destruction.

As noted in Section 3.2, most distributed burners can be axially fired into an unbaffled combustion chamber and mixing will be fast enough to give a nearly uniform temperature profile in 10' - 20'. However, with discrete burners simple turbulent flow cross mixing and in most cases axial jet mixing cannot be expected to bring all fume up to the required incineration temperature in the distance available in commercial afterburners. Burner location, baffles, and combustion chamber configuration are all used to improve mixing and afterburner performance. Chapter 14 should be consulted for a more detailed discussion of various mixing arrangements and a discussion of what constitutes adequate mixing in practical afterburners.

When Maxon was first applying their Combustifume distributed burner (Section 3.2.1.1) they compared its performance with that of a single axial fired burner in a ~4' diameter unbaffled combustion chamber with length ~2 - 3 times the diameter. Fume was preheated in a recuperative exchanger. Results showed that outlet hydrocarbon concentration was reduced to ~10 ppm at 1250°F with the distributed burner. Concentrations were >200 ppm using the discrete burner, and the nominal combustion chamber temperature had to be raised above 1550°F to achieve similar hydrocarbon destruction. In addition, the combustion chamber length had been reduced in installing the distributed burner. Since hydrocarbon destruction is rapid at 1100° - 1200°F (Section 3.1.4.2) this demonstrates the problems caused by poor mixing.

As discussed in Section 3.2.3.2, cross sectional area should be chosen so that flow will be strongly turbulent so that mixing occurs rapidly on the molecular level to heat and oxidize all pollutants. This is important for small scale mixing. However, additional means must be provided for cross channel mixing. In Section 3.2.3.2 it was noted that mixing is improved by high velocities and high length to diameter ratios. This is limited by pressure drop across the afterburner and mechanical design considerations.

3.3.2.1 Axial Flame-Jet Mixing

A common afterburner arrangement is to utilize a single burner firing into the end of the combustion chamber with fume brought in around the burner as a parallel stream (see Figure 3-12). If a high excess air burner (Section 13.2.6.6) is used, the fume may all go through the burner but part of the fume (~50%) must still flow around the flame and mix in downstream. As noted in Section 14.4, surrounding fume will slowly be entrained into the hot combustion gases so there is little danger of flame quenching. However, large temperature variations would be expected across the exit end of an unbaffled combustion chamber with length to diameter ratio less than 10 - 15. Typical length to diameter ratios found in available afterburners are ~2 - 6 so much fume can pass around the hot combustion gases and escape without being raised to a sufficiently high temperature. Additional mixing promotors are required.

One means of improving simple axial jet mixing is to reduce the diameter of the combustion chamber after fume has been entrained into the jet of hot combustion gases. One means of doing this is shown in Figure 3-17. This particular fume incinerator utilizes the flame jet to create an eductor effect which pulls fume into the afterburner. Only that fume which can be entrained by the jet is pulled in. If instead, the fume were to be supplied under pressure, it would not be completely entrained and cold fume would surround the hot gases as they flow through the throat. The situation would be similar to that discussed above. In either case there will be considerable flame quenching unless the flame extends only a short distance beyond the burner block. Addition discussion can be found in Section 14.3.3.

Some early afterburner designs employed a discrete burner firing upstream into the fume flow. This may result in some increase in mixing rate but the amount of flame quenching was disasterous. Flame stability was also a problem when fume flow rate was high. This arrangement should be avoided, unless fuel is burned in a separate chamber and only hot combustion products mix with the cold fumes.

3.3.2.2 Radial Entry of Fume or Flame

Cross stream mixing is much faster when the streams to be mixed enter at right angles (or at least some angle) rather than parallel. Burners often enter from the side, especially in stack afterburners but this usually leads to flame impingement and shortened refractory life (See Section 7.2). Also the improved mixing and attempts in many designs to "fill the cross section with flame" lead to severe flame quenching and formation of CO and aldehydes (Section 3.1.4). In a high temperature (~1400°F) afterburner such flame quenching is not too serious (if good downstream uniformity exists) since intermediates from the fuel will be oxidized with the fume. However, in a low temperature (~1000°F) smoke or odor abator the CO and aldehydes will remain in the flue gas and cause an increased pollution problem. If low temperature units are to be used, fuel should be burned in a separate chamber (burner block extension) and only hot combustion products should mix with excess fume.

A more satisfactory arrangement employes an axially fired burner with multiple, high velocity cross flow jets of fume mixed in downstream of the flame.

More typically the fume enters through one or two radial streams at relatively low velocity and does not penetrate the flame jet very far. This is similar to the multijet burner in Figure 3-7b. Baffles must be provided to enhance downstream mixing since the situation is similar to axial entry of both fume and flame.

3.3.2.3 Tangential Entry of Fume or Flame - Swirl

As discussed in Section 3.3.1.6 swirl can be used in burners to create intense mixing and give very short flames. Tangential entry in an afterburner can also be used to speed up mixing of bypassed fume and hot combustion products. Pressure drop must be expended to generate swirl and speed up mixing. In vortex burners this pressure drop may be ~18" w.c., but typically the total pressure drop across an afterburner is only 2" - 4" w.c. so relatively slow mixing is expected.

The cold fume stream should enter axially and the hot combustion products should enter tangentially for improved mixing performance. If instead the fume is brought in tangentially at high velocity, the cool (and therefore heavier) gases remain near the wall due to cyclonic action and mix more slowly with the flame than if both were brought in axially. Tangential firing of the burners is preferred. A vertical upflow afterburner with tangential firing is a good arrangement for incinerating fumes containing heavy particulates (Section 3.1.2.6) since they are held near the bottom of the combustion chamber until consumed and experience a much longer residence time than the average.

Flame quenching will be a problem with tangential flame entry, especially in designs which attempt to fill the combustion chamber with flame. (In most cases the flame actually follows a helical pattern along the wall of the afterburner since the flame jet is weak compared to the fume stream, especially at lower firing rates.) As noted in the previous section a certain amount of flame quenching can be tolerated in 1400°F afterburners with uniform flow downstream since these generated pollutants will also be oxidized. However, it would be better to complete the flame within a burner extention and inject the combustion products for mixing with fume. Refractory used with tangential flame entry will be subject to flame impingement and possibly require increased maintenance.

As in the swirl vortex burners (Section 3.3.1.6), mixing will be improved if the combustion chamber diameter expands immediately downstream of the point where swirling flow is generated. This generates backflow and much cross chamber mixing. Where vortex burners are used, some swirl is still present in the combustion gases issuing from the burner and can be used to provide some energy for mixing with the bypassed fume.

3.3.2.4 Baffles

In most discrete burner fume incinerators on the market today, the combustion chamber is not long enough to provide adequate mixing (and pollutant destruction) without the use of baffles. Baffles can greatly speed cross chamber mixing of fume and combustion products. However, they must be used correctly or else they will only add pressure drop and set up recirculating dead zones in their wake which remove active chamber volume and shorten effective residence time without improving mixing. Section 14.8 contains a more detailed discussion of baffles.

Baffles can be used to compensate for lack of a distributed source of hot combustion gases by diverting streams over relatively large distances across the combustion chamber. Also they serve to block off a major part of the chamber cross section so flow velocities are rapidly increased and then decreased on the downstream side. This generates a high level of turbulence and speeds local mixing at the expense of pressure drop.

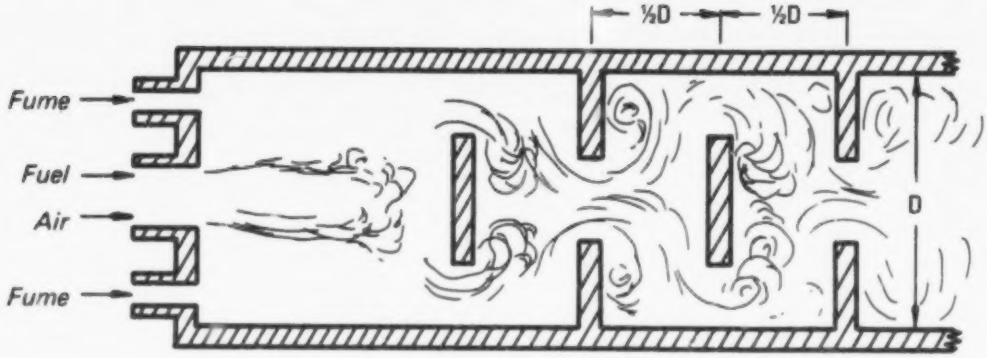
As discussed in Sections 14.8.1 and 14.8.2 ring and disc baffles and bridge wall baffles, shown in Figure 3-18a and 3-18b, are effective in forcing cross channel mixing over large distances. They should be used in pairs for best results and sized to block 60% or more of the chamber cross section. This will add several inches of water pressure drop across the afterburner, but will improve performance. Fairly good mixing has been demonstrated with ring and disc baffles at the same point in the afterburner so that an annular slot is formed with ~25% free area (See Figure 3-12). As discussed in Section 14.8.2 less than 50°F temperature variation is obtained within 2 chamber diameters downstream of this baffle.

Baffles should be located so as to complete mixing as rapidly as possible leaving maximum residence time after all fume is brought to temperature. Flame impingement is to be avoided so the first baffle is usually 1' - 2' beyond the maximum burner flame length. Subsequent baffles should be located ~1 chamber diameter downstream, similar to what is shown in Figure 3-18. The common placement at 1/3 and 2/3 of the chamber length is probably not optimum.

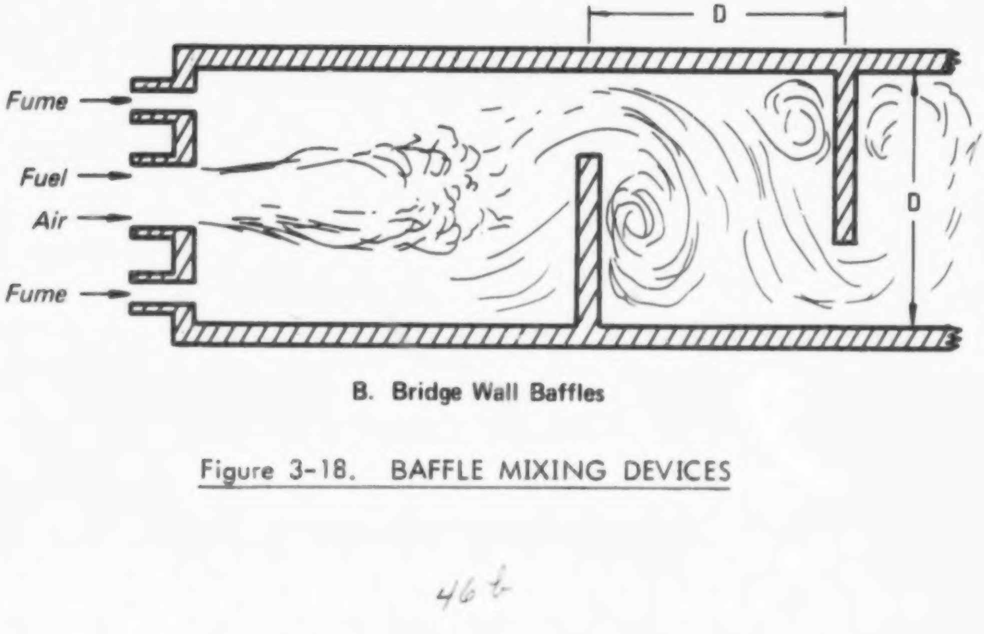
Section 3.2.3.3 contains a discussion of the important role ring baffles and grids can play in afterburners utilizing distributed burners. Ring baffles along the wall tend to prevent escape of cold fume from the afterburner by flowing along a cool wall. Such wall flow is diverted into the main stream. These would also play the same role in otherwise unbaffled afterburners utilizing discrete burners. However, overall performance will be much better if gross cross channel mixing is provided by the disc or bridge wall baffles just discussed.

The small scale of cross stream mixing provided by grids or checkerwalls (bricks are layed so as to form a gridwork across the chamber) also limits their usefulness with discrete burners. The gridwork can serve to eliminate velocity gradients if a pressure drop ~1" w.c. is taken over the grid but temperature and concentration non-uniformity will only be affected over the scale of the gridwork (~6"). In one case where three small discrete burners were equally spaced across the entrance to a 1' by 5' cross section combustion chamber, a checkerwork grid was put across the chamber in an attempt to improve mixing and increase the <70% efficiency of pollutant destruction. The overall destruction efficiency was increased only ~2% because the scale of maldistribution was much greater than the size of the gridwork and fume continued to bypass without reaching incineration temperature. (Heat exchanger leakage also contributed to poor performance.)

Figures 3-18a, b thru 3-20 follow


½D
½D
Fume
Fuel
Air
Fume
D

A. Ring and Disc Baffles


D
Fume
Fuel
Air
Fume
D

B. Bridge Wall Baffles

Figure 3-18. BAFFLE MIXING DEVICES

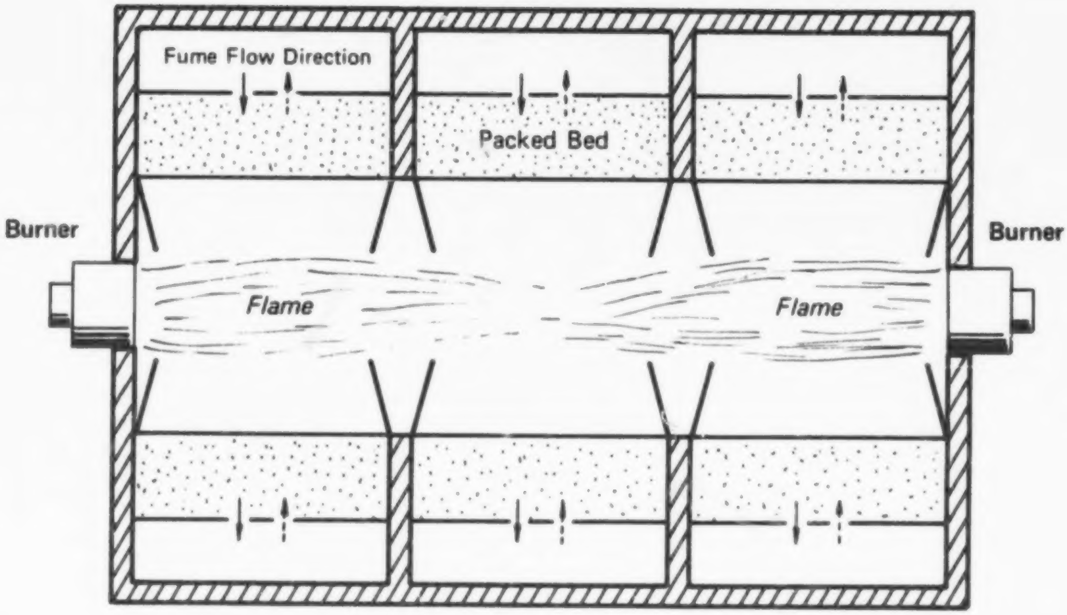

Fume Flow Direction
Packed Bed
Burner
Flame
Flame
Burner

Figure 3-19. PROCTOR-SCHWARTZ TRAPS SYSTEM WITH REGENERATIVE HEAT RECOVERY
Top View with Ducting Removed

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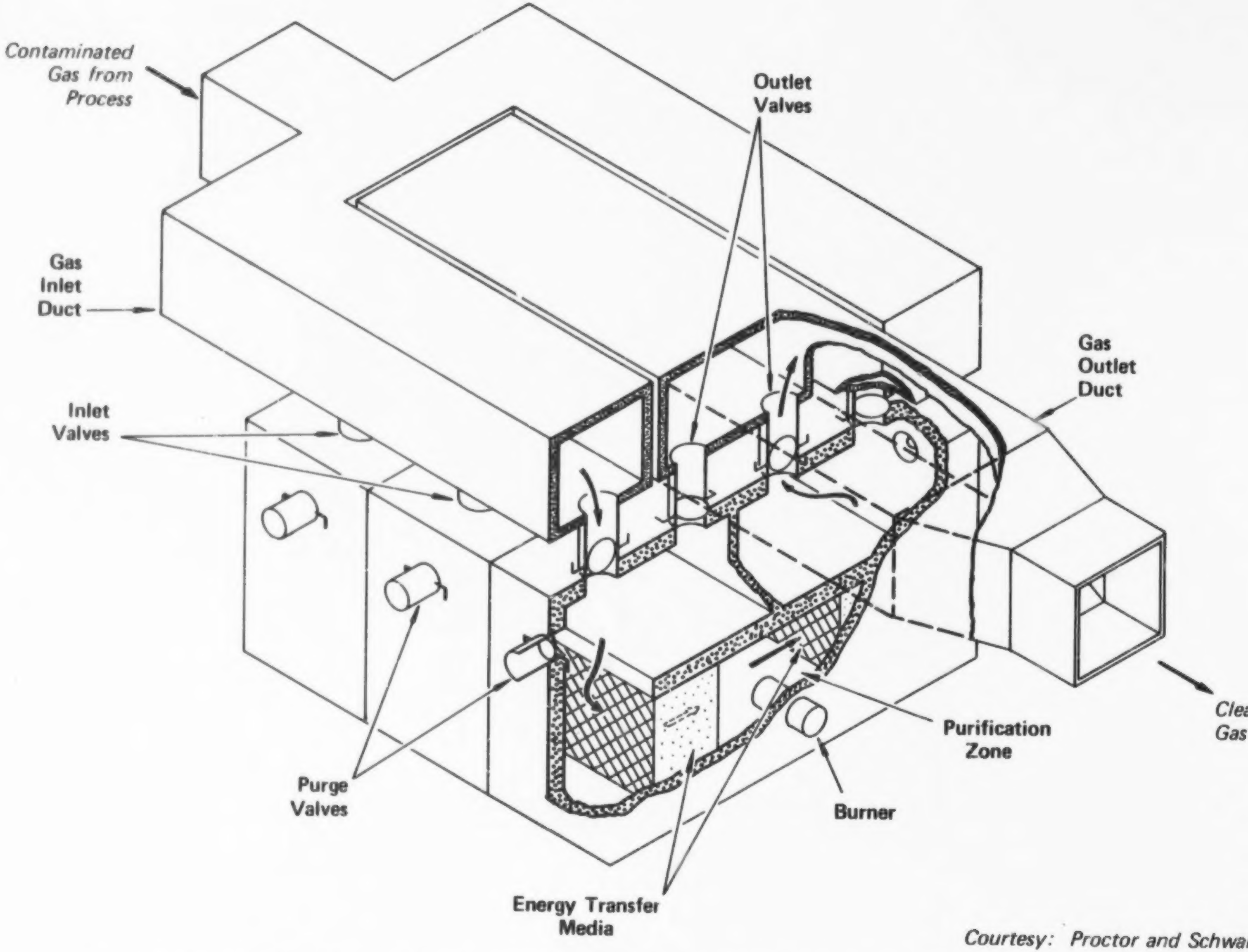

Contaminated
Gas from
Process
Outlet
Valves
Gas
Inlet
Duct
Gas
Outlet
Duct
Inlet
Valves
Clean
Gas
Purification
Zone
Purge
Valves
Burner
Energy Transfer
Media

Courtesy: Proctor and Schwartz, Inc.

Figure 3-20. VALVE AND DUCTING ARRANGEMENT IN PROCTOR SCHWARTZ TRAPS SYSTEM

3.3.2.5 Bends and Changes in Combustion Chamber Cross Section

As is discussed in Section 3.4.1 flow around a 90° or 180° bend is not a good mixing promotor unless the combustion chamber extends for ~6 - 10 chamber diameters after the bend rather than the normal 2 diameters or less. Recirculating dead zones are set up which reduce the total residence time in the chamber so such bends should be avoided where possible.

If strong swirling flows are involved, a sudden expansion of the combustion chamber cross section can set up recirculation which significantly speeds mixing. With simple axial flow abrupt expansions and contractions of the cross sectional area produce some increase in mixing rate, but this is more effectively done using baffles.

3.3.2.6 Packed Bed Regenerative Heat Exchange - TRAPS System

Heat recovery systems and associated heat exchangers are discussed in Chapter 6 since they are usually add on devices. They make little difference in the operation of the combustion chamber except that a hotter inlet fume must be handled and little mixing energy will be available from the small supplemental fuel flame required. This is not the case with a new high heat recovery fume incineration system offered by Proctor Schwartz under the name TRAPS. A sketch of this system (top view) is shown in Figure 3-19. The unit supplied for a minimum fume flow ~10,000 SCFM is composed of a minimum of three modules (as shown in Figure 3-19) with packed bed heat exchangers along the sides of the combustion chamber and a discrete burner firing from each end. Fume flows in through a packed bed exchanger where it is preheated to ~1100°F, flows across the combustion chamber where it is raised to 1400° - 1500°F, and flows out through the packed bed exchanger on the opposite side where it is cooled to ~600°F. Ductwork and butterfly valves are arranged to reverse the flow direction in each module every ~2 minutes so the cold incoming fume can pick up the heat transferred to the packed bed exchanger by the hot exhaust fume on the previous half cycle. Figure 3-20 indicates the arrangement of these valves and ducting. Between each half cycle a short air purge is required to sweep contaminated fumes out of the region between the inlet valve and the combustion chamber. Otherwise, when the flow direction is reversed, this fume would be swept out the exhaust without treatment.

The regenerative packed bed exchangers offer high heat recovery efficiency; the manufacturer claims 75% or 83% recovery depending on the depth of packed bed used. The ceramic packing has a high heat capacity so only 25° - 50°F temperature variations are expected at any point in the packed bed during the ~2 minute half cycle. (There is a sharp temperature change across the bed, ~1300° on the combustion chamber side and ~500° on the inlet (outlet) side.) Pressure drop (~8" w.c.) is expended in obtaining this high heat recovery efficiency.

In order to minimize pressure surges in the fume inlet duct (which might upset the fume generating process) each module is operated on an independent cycle so that only one set of valves is closed for air purging at any time. Therefore, adjacent modules will often have flow in opposite directions, and there will be a strong tendency for fume to enter the combustion chamber through one packed bed and flow out through the adjacent packed bed without crossing the chamber. This could lead to very short residence times in the chamber compared to that calculated for flow across and out the opposite side and necessitates the use of the baffles shown in Figure 3-19 between each module. These baffles insure a minimum flow path between modules of .3 - .4 seconds residence time at 1400°F. Some pollutant destruction probably also occurs within the inlet packed bed.

Problems could arise in achieving good pollutant destruction if large temperature gradients can exist within the combustion chamber. Cool flow paths could then be set up and stabilized by the heat exchange process. As in conventional designs, the control thermocouple only records the temperature at a single point. Mixing within the ~8' x 5.5' x 11' (for three modules) combustion chamber is provided by the flame jets at each end. Their velocity is several times higher than the nominal cross flow velocity of the fume but relatively little heat input is needed (~2 - 3 MM Btu/hr from each burner) because of the high heat recovery. For fume flow rates above 10,000 SCFM more modules are added but two burners continue to be used and must provide the mixing action.

As mentioned previously these units are new and operating experience was not available for us to assess the adequacy of chamber mixing and therefore pollutant destruction. Such experience is also needed to determine the ease of operating on fouling fume streams. Organics should be able to be removed from the packed beds by burning off, but inorganics may require repacking of the exchangers.

3.4 Thermal Afterburner Design Features - System Configuration and Construction

In this section we consider thermal afterburner design features which are independent of the type of burner used. The combustion chamber must be designed to provide uniform treatment of all fume once it is raised to the incineration temperature and hold it for sufficient time for oxidation to occur (Section 3.1.2). Operation can also be affected by the orientation of the unit, especially when mounted vertically to serve as its own stack. In addition to providing pollutant destruction, the afterburner must be constructed with sufficient mechanical strength to withstand the 1400° - 1500°F combustion chamber temperature and rapid heating and cooling cycles.

3.4.1 Plug Flow Retention Section - Straight Through Circular Chamber

Once all of the fume has been raised to the required incineration temperature a plug flow condition should be maintained so that all of the fume is given equal residence time at this temperature. (Plug flow is an idealized picture in which fluid moves as a solid and there is no backmixing or cross mixing between different locations in the gas.) When an approximation to plug flow is not maintained, large variations in velocity and recirculation patterns can exist. This means that some of the fume can be swept through the chamber in a small fraction of the calculated residence time while others remain trapped for a long time in recirculating "dead zones." Of course plug flow should be avoided until all fume has been mixed with the hot combustion products. Otherwise temperature nonuniformity will be preserved in the short flow path provided in available afterburners and fume will escape without sufficient pollutant destruction.

As is discussed in Chapter 15 a straight, unbaffled chamber with circular cross section will give a close approximation to plug flow and residence time will be the chamber volume divided by the actual volumetric flow rate. Cross stream mixing is very slow so nearly uniform temperatures must be obtained by mixing upstream. Afterburners commonly have a rectangular cross section because of ease of fabrication, especially for large sizes. Some circulation occurs in the corners and flow may be restricted in these regions. As a result residence time is slightly less than calculated from the total chamber volume.

More serious reduction in residence time can occur with 90° or 180° turns in the combustion chamber. The 180° or U bend construction in which the combustion chamber is folded back on itself is quite common in available afterburners. It achieves a slight cost reduction and a considerable reduction in total length since less exterior wall is used. Also, if recuperative heat recovery is desired, it is easy since inlet and outlet are close together. However, some problems have been experienced with overheating of the dividing plate since it is heated from both sides. Materials selection, either metal or refractory, is crucial (Chapter 7). We encountered several cases in which this baffle warped, collapsed, or required extensive repair.

The 180° turn is usually put forth as a mixing promoter which will improve afterburner performance, but instead it more likely results in less favorable performance. Cross stream mixing is improved downstream of such a bend but 6 - 8 combustion chamber diameters are required to achieve such mixing. Since afterburners typically have L/D ratios ~2 - 6 with only half of this following the 180° bend, not much mixing will be accomplished. Figure 3-21a shows the recirculation patterns which are likely to develop downstream of the U bend and in the corners of the chamber. These are fairly stable patterns and can result in as much as 25% of the chamber volume being unavailable for fume flow. An equivalent reduction in residence time results. As discussed in Chapter 15 large variations in velocity are found within the active chamber volume so that some fume will be swept through the chamber in a small fraction of the average residence time while other parts of the fume recirculate for a long time before being released. Figure 3-21 also shows the recirculation patterns developed downstream of 90° turns.

A checkerwork baffle or grid should be placed at the start of the retention section if mixing patterns can be determined well enough to define the point at which fume has been sufficiently well mixed with hot combustion gases. This grid will serve to even out velocity variations across the chamber and generate turbulence on a scale which speeds mixing and reaction on the molecular level.

3.4.2 Vertical or Stack Afterburners

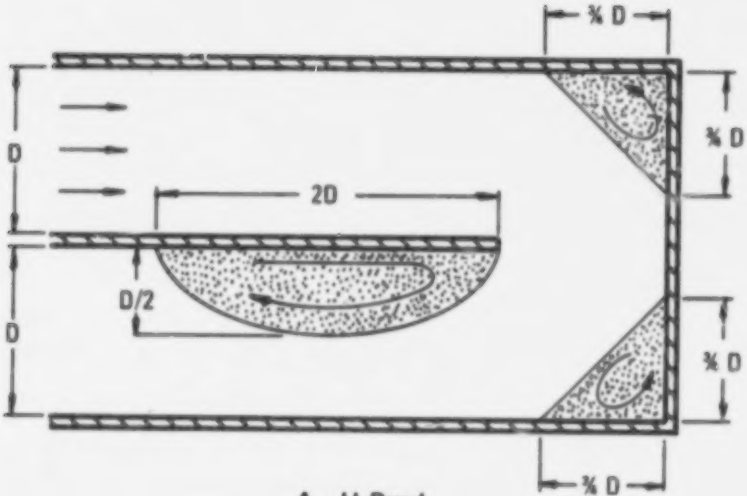
Considerations specific to each installation usually determine whether a vertical or horizontal afterburner will be used (location, allowable roof loading, expected wind speeds, etc.). Gravity can cause some stratification between hot combustion products and cold fume and make mixing more difficult in a horizontal unit, but this is a minor consideration since flow velocities are high and buoyancy effects are usually small. Vertical units serve as at least part of their own exhaust stack which is an advantage but also can lead to problems. If the vertical unit has an open top and has a short length to diameter ratio, much heat will be lost from the combustion chamber by radiation to the sky. Insufficient treatment of much of the pollutant can result. Also strong cross winds may affect flow patterns within the chamber and result in cold zones. Both of these problems make it difficult to find a suitable location for the control thermocouple.

In order to avoid these problems, a vertical afterburner combustion chamber should have a reduced cross section at the exit end to give a higher exit velocity and reflect much of the radiation which could be lost. Also a rain hat can serve to reduce such losses. Short, squat vertical fume incinerators (L/D 2 - 3) should be avoided and a stack extension would be recommended. Horizontal afterburners require a 90° turn into the stack so these problems do not occur.

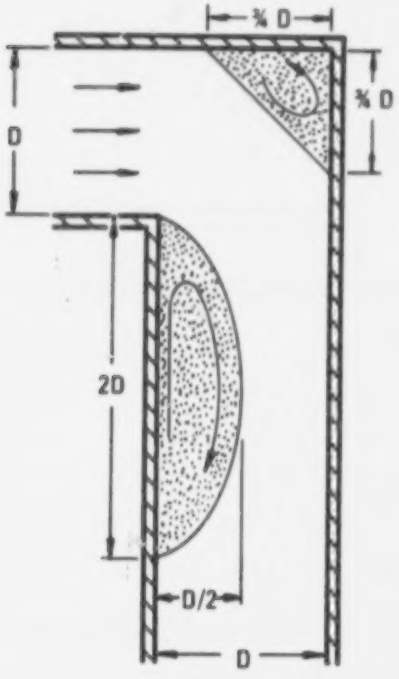
3.4.3 Induced Draft - Avoiding Leakage

Combustion chambers can be run at pressures either above or below atmospheric pressure by placing the fume blower either before the afterburner (forced draft - positive pressure) or following the afterburner (induced draft - negative pressure). If leaks develop in a chamber under slight negative pressure, outside air will enter and produce a very small increase in fuel requirement.

Figures 3-21, 3-22 follow


¾ D
D
¾ D
2D
D/2
D
¾ D
¾ D

A. U Bend


¾ D
¾ D
D
2D
D/2
D

Shaded Areas Represent
Recirculating Flows –
Dead Zones

B. L Bend

Figure 3-21. REDUCED RESIDENCE TIME EFFECT OF 180° AND 90° TURNS

50h

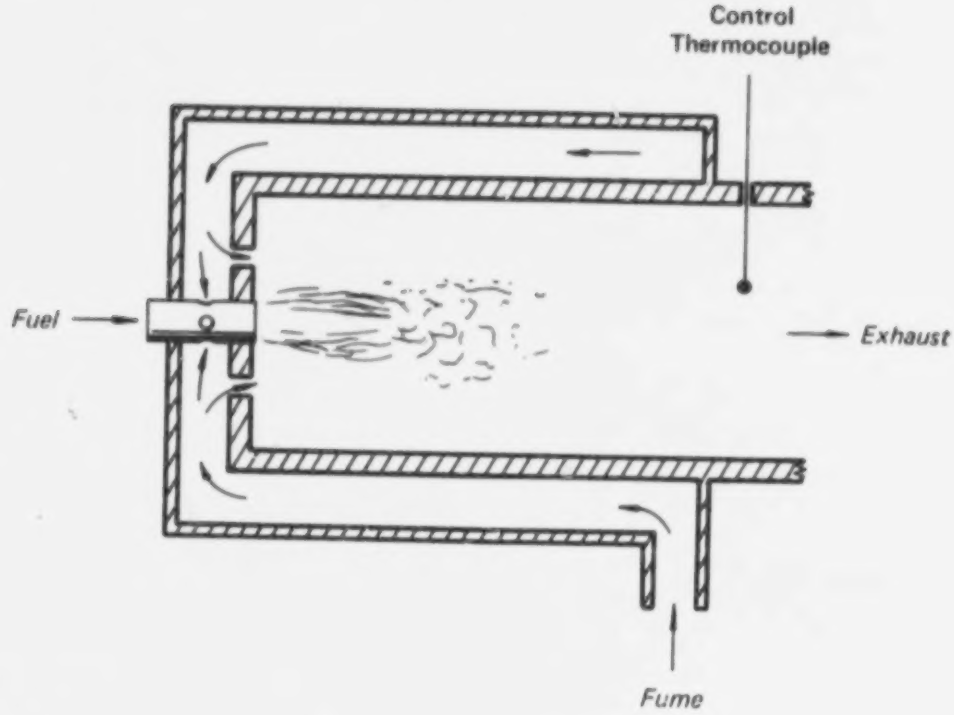

Control Thermocouple
Fuel
Exhaust
Fume

Figure 3-22. ALL METAL AFTERBURNER - ANNULAR DESIGN

However, in a forced draft system hot gases will escape through the leak and will probably make it bigger. The gases issuing from the leak probably still contain pollutant and may be toxic and offer danger to nearby personnel because of their high temperature. Through proper mechanical design and use of heavy gauge metal and careful assembly the possibility of leaks can be minimized. Induced draft systems offer an extra degree of safety, but higher blower costs are incurred (see Section 8.2). Some afterburners designed for forced draft operation, are constructed with a double metal wall around the refractory. A separate, low volume blower delivers outside air to the space between the walls at a pressure greater than that in the combustion chamber. This again prevents fume and hot gases from escaping and allows use of loose expansion joints to prevent thermal stress buildup.

Many problems have occurred with tube leaks in recuperative heat exchangers due to thermal stresses and improper tube fastening. With either forced or induced draft systems the clean, hot flue gas will be at lower pressure than the polluted, cold fume. Therefore, untreated fume leaks into the cleaned exhaust gases and greatly reduces the efficiency of the afterburner system. The only way the leakage direction can be reversed is to locate the blower between the preheater and the combustion chamber, not a common practice.

3.4.4 All Metal Construction - Annular Fume Inlet

Many early afterburners were essentially burners in a stack designed to eliminate a visible plume and operate at 600° - 800°F. At these temperatures carbon steel or low alloys could be used successfully. However, as noted in Section 3.1.2.5 and elsewhere, such low temperature afterburners cause more pollution than they abate and should not be used. High temperature (~1400°F) afterburners should be constructed with a refractory lining to act as an insulator and protect the outer metal shell which provides structural strength and seals the unit. Refractory selection, cost, and installation techniques are discussed in Section 7.2.

Refractory is heavy, adds expense to the unit, and requires relatively slow warm up of the afterburner. (Refractory can be damaged if heated or cooled too rapidly, but it also has a high heat capacity so a large refractory lined afterburner will take a minimum of 5 - 15 minutes to come up to operating temperature from a cold start.) For these reasons all metal construction is used in some units. The most common design involves a double shell as indicated in Figure 3-22. Cold fume comes in around the combustion chamber and keeps the metal wall cool.

Our survey indicates that this design has led to a great amount of trouble, primarily structural failure. This has resulted in severe leakage, the need for extensive repairs, and in some cases abandonment of the afterburner as inoperable. The problems seem linked to the inability to handle thermal stresses which are developed due to the sharp temperature gradients in this design. In some cases there has been evidence of excessive temperatures due to combustion of condensate which built up in the annular space.

It has also been claimed that this annular design serves to preheat the incoming fume and recover heat, but this is only marginally true if at all. The amount of heat transfer area is very limited so even if fins are used to increase heat transfer, only 100° - 200°F preheat will be experienced by the fume. But in the arrangement indicated in Figure 3-22 even this amount of recovered heat cannot be utilized. The heat needed to preheat the fume comes from the gases in the combustion chamber so to hold 1400°F at the chamber exit the average temperature at flame end must be 1500° - 1600°F. The same amount of fuel must be burned as if the fume entered without preheat and a 1400°F exit temperature is maintained.

Perhaps the structural failure could be overcome through proper mechanical design and use of heavy gauge stainless steels. However, the design still suffers from the presence of cold walls in the combustion chamber. As discussed in Sections 3.2 and 3.3 wall flow leading to fume escaping without adequate treatment is a potential problem with refractory lined fume incinerators. If the walls are "cold," quenching of the oxidation reactions will occur and fume escape is nearly inevitable.

3.4.5 Afterburner Should Be Rugged and Well Designed Mechanically

Many light weight, poorly constructed, poorly controlled afterburners have been sold, usually at a low price often by companies anxious to get into the pollution control field. These units are often bought by companies anxious to satisfy some new air pollution code or to quiet complaining neighbors. In many cases structural failure results in a year or less and the money spent is lost or much money and effort is spent trying to beef up the unit.

Mechanical design of afterburners is beyond the scope of this handbook, but it is important that one deal with an experienced and reputable manufacturer. A major difficulty is involved with handling the thermal expansion which occurs on heating the unit to 1400° - 1500°F and even hotter in the vicinity of the flame. A long unit may expand several inches so provision must be made to allow for this movement and minimize thermal stresses. Such stresses are a special problem when hot and cold gases are close together (as in a heat exchanger) or a surface is rapidly heated and cooled. Problems also arise when materials with vastly different thermal expansion coefficients (metal and refractory) are used together. Chapters 6 and 7 contain additional discussion of methods of avoiding or minimizing thermal stress buildup.

In most units mechanical strength and sealing are provided by an exterior metal shell. As discussed in Chapter 7 most steels, even stainless steels, lose much of their strength at temperatures of 1000°F and above. Therefore a refractory lining is used to keep the metal temperature much lower. The advantages and drawbacks of various metals and refractories are discussed in Chapter 7. Afterburners are manufactured with shell thicknesses ranging from 20 gauge to 3/8" plate. In general, the thin metal shells have not had sufficient mechanical strength to stand up under typical afterburner service.

3.5 Users Survey on Thermal Afterburner Performance

Questionnaires were filled in covering 214 thermal afterburners. A complete tabulation of the data is given in Chapter 11 (Table 11-2). Pertinent information has been extracted and summarized in Table 3-5. In general, the thermal units were reported to be satisfactory overall, only 11% of the units being regarded as unsatisfactory for any cause.

The percentage reduction in hydrocarbon was given in only a limited number of cases, and it was difficult to determine the reliability of the figures quoted. Most units were claimed to give greater than 90% reduction where numbers were reported. Generally units are adjusted during start-up to meet the local pollution requirements, a temperature is determined at which satisfactory operation is obtained, and no further testing is carried out. If the unit is operating at the temperature agreed to by the local control board, then it is assumed to be doing an adequate job.

Main operating problems involve safety controls, refractory linings, heat exchanger fouling or mechanical failure, and bearing failure in fans. However, most units are available 95-99% of the required operating time, so that maintenance is a minor problem.

Table 3-5. USERS INFORMATION ON THERMAL AFTERBURNER PERFORMANCE

Overall Performance

Number Installed	Number with Heat Exchangers	Number Not Yet Run	Performance		
			No Comment	Satisfactory	
				Number	% of Those in Use
214	83	7	2	184	89

Reported Hydrocarbon Reduction

% Reduction in HC	95-100	90-95	85-90	80-85	Total
Number reported as meeting requirements	41	21	2	2	66
Number not meeting requirements			1	1	2

Main Operating Problems

Category	Safety Controls Flame-Rod	Refractory	Heat Exchanger	Fan
Number of afterburners having problems in each category	32	19	16	10

3.6 Conditions Required for Adequate Performance in Some Applications

Table 3-6 presents operating temperatures recommended or used for thermal afterburner applications in various industries. The first column gives temperatures recommended by the LAAPCD (taken from Table 51 of the revised edition of reference 3-1) for these applications when using afterburners having 0.3-0.5 seconds residence time. The second column presents the range of temperatures reported by users (in the Users Survey) to give generally satisfactory operation. The final column gives temperatures which have been quoted in various literature articles describing the use of thermal afterburners for these applications.

Conditions in the usual paint baking and drying oven are such that the oven burner does not contribute significantly to the destruction of oven fumes. Claims have been made by the manufacturer of one oven design that its use of air-jet-driven recirculation and mixing of the oven atmosphere with the hot gases from the burner results in significant oxidation of fume material in the oven. These claims are not substantiated by available test data. It is possible that some material may be oxidized, but only a small fraction of the oven contents will be heated to a temperature in excess of 1300 or 1400°F for a tenth of a second or so, as is required to achieve a significant conversion. The rest of the oven gases can be expected to remain at a well-mixed temperature near the normal 300 to 500°F oven operating temperature and remain unoxidized.

3.7 References

Many articles on afterburners have appeared in trade and technical journals, but few are referenced here. Most of these articles are in the nature of advertising and discuss the features of a particular manufacturer's unit. Articles discussing necessary design criteria and types of applications are much less complete and objective than the material contained in this handbook. In many cases essentially the same material has been republished time and again. A few articles are thought to be useful because they relate experience with afterburner installations in particular industries and contain some performance data. However, as noted previously, most performance data is not useful since the operating conditions are not constant and the design features of the unit are often not reported.

The "Air Pollution Engineering Manual" prepared by the Los Angeles Co. Air Pollution Control District[3-1] is recommended for its industry by industry treatment of the air pollution problem and control techniques. Other useful references to supplement this handbook are 3-2 to 3-7 found in the Bibliography.

Table 3-6 follows

Table 3-6. TYPICAL THERMAL AFTERBURNER OPERATING TEMPERATURES (°F)

Industry	LAAPCD Recommendation[a]	Survey Data	Literature
Asphalt blowing		1000-2000	
Biological control, fermentation		1100-1250	
Carpet laminating	1200-1400		
Coffee roasting		1200-1500	1050
Coil coating, sheet coating, metal decorating	-	1200-1500	1300
Core ovens, foundry	1400	2000	
Coating, engraving		1000[b]-1450	
Cloth carbonization	1800		
Deep fat fryers	1200		
Gum label drying oven			1250
Mineral wool, fibreglass curing		1000[b]	1310
Odor control (general) sludge off-gas	1300-1500	1100-1250	1300-1425
Hardborad tempering		1200	1200
Oil and grease smoke (metal chip recovery, heat quench baths, tempering)	1200-1400	900[b]-1600	
Paint bake ovens	1200-1500	1100-1500	1240
Paper manufacture - sulfite digester off-gas			1350
Pipe wrapping	1400		
Rendering plants	1200	1200-1300	1200
Rubber products		1300-1400	1300
Petroleum refining and products		1300-2000	
Printing, lithographing		1300-1500	
Smelting, refining, metal recovery, wire burnoff		1300-1650	
Smokehouse operation	1200	800[b]-1200	
Solvent control	1300-1500		
Varnish cookers, resin kettles	1200	1000[b]-1500	1200-1400
Vinyl plastisol curing	1200-1400	1200	900[b]
Wood milling			1200
Wire enameling	-		1300-1350, 1400
Phthalic anhdyride		1250-1440	
Textile drying oven		1350	

a) Reference 3-1, revised version, Table 51.
b) Low temperature generally for odor, smoke control, not true fume destruction.

Chapter 4. CATALYTIC AFTERBURNERS: DESCRIPTION AND PERFORMANCE

4.1 Introduction

A catalytic afterburner is an alternative to a thermal incinerator as a means for oxidizing gaseous hydrocarbons (or oxygenated hydrocarbons) to carbon dioxide and water. Contact of a waste stream with a catalyst bed (or catalyst matrix structure) allows oxidation reactions to occur rapidly in the temperature range 700 - 900°F, in contrast to the 1300 - 1500°F required for practical oxidation rates in thermal afterburners. The catalyst conforms to definition by accelerating the oxidation reactions without itself being changed in any way. The oxidation reactions which occur at the surface of the heterogeneous catalyst produce the same products: carbon dioxide and water and liberate the same heat of combustion as does thermal oxidation.

4.1.1 Basic Elements of System

The basic elements of a catalytic afterburner system are shown schematically in Figure 4-1. Heat necessary to bring the waste stream up to the required (catalytic) oxidation temperature is normally supplied by a preheat burner. A combustion/mixing chamber downstream of this burner is designed to achieve a uniformly preheated, uniformly distributed mixture of combustion products and waste stream, which is then passed through the catalyst bed. The catalyst bed (or matrix) in commercial units is typically a metal mesh-mat, ceramic honeycomb, or other ceramic matrix structure with a surface deposit or coating of finely divided platinum or other platinum family metals. The finely-divided metal deposit is the actual catalyst for the oxidation reactions, while the matrix serves to support the catalyst on a high geometric surface area and promote good contact between waste stream and catalyst. A surprisingly small volume of catalyst bed is required, typically 0.5 - 2 ft^3 catalyst/M SCFM waste stream for 85 - 95% conversion of hydrocarbons in the waste stream. The small volume and the low density of the catalyst contribute to overall small size and light weight for catalytic units. Recovery of heat from the cleaned effluent stream may be included in the system using any of the techniques used for this purpose with thermal afterburners (e.g., recuperative exchange with the cold waste stream, or recycle of part of the hot, cleaned stream to a drying oven). However, because required preheat temperatures are lower and hence fuel consumption is lower, there is often less economic incentive for inclusion of heat recovery in a catalytic afterburner than in a thermal afterburner system.

4.1.2 Economics Relative to Thermal Afterburners

The comparison of equipment costs for catalytic and thermal afterburners depends somewhat on the capacity of the unit and type of system. For small capacity (up to 10 - 12 M SCFM), pre-engineered catalytic units, the purchase cost is substantially equal to that of a thermal afterburner of equivalent capacity (see Chapter 11). For larger, custom designed units, purchase costs for catalytic afterburners are somewhat higher than for thermal units. The inclusion of a single pass, recuperative heat exchanger in both systems again gives approximately equal purchase costs for catalytic and thermal systems since the heat exchanger for the catalytic system handles lower temperatures and may be constructed using less expensive materials. Because catalytic systems are frequently smaller and lighter than corresponding thermal systems, the cost and convenience of installation usually favors catalytic systems.

The fuel consumption for preheating in a catalytic system is typically 40 - 60% of that required by a corresponding thermal afterburner. This economy in fuel consumption is partially offset by increased operating costs for maintenance (periodic cleaning of catalyst), and for periodic replacement of the catalyst element in the system. Although the catalyst is unchanged by the oxidation reactions which it promotes, it is subject to more or less gradual decrease in effectiveness due to three factors. 1) Even "particulate free" waste streams carry low concentrations of particulates (usually inorganic materials) which gradually build up a deposit on the surface of the catalyst, restricting contact between waste stream and active catalyst, reducing performance of the unit. Catalyst manufacturers recommend a number of physical and chemical cleaning methods for removal of such surface deposits. With such periodic cleaning at 3 - 12-month intervals, it is normally possible to maintain effective catalyst activity at a satisfactory level. 2) With continued exposure to operating conditions, an irreversible decline in catalyst activity takes place due to microstructural changes in the active material in the catalyst matrix. Under normally recommended conditions, catalyst activity loss due to this factor would require replacement (or reactivation by the manufacturer) of the catalyst element at 3 - 5-year intervals. Continued exposure of the catalyst to higher than normal temperature accelerates this aging-deactivation process and shortens the effective life of the catalyst element. 3) The presence of specific catalyst poisons (e.g., zinc, phosphates--see discussion in section 4.2.4.3) in the waste stream being treated will also shorten the effective life of the catalyst. If the combination of factors contributing to catalyst deactivation allow a catalyst life (between replacements or reactivation by the manufacture) of one year or longer, a catalytic afterburner will normally be more economical than a thermal afterburner. Manufacturers of catalytic afterburner systems (who all also manufacture thermal afterburner systems) will not generally recommend use of a catalytic system for an application where catalyst life of less than one year is anticipated.

A summary of comparative costs associated with thermal and catalytic afterburners in a typical installation is presented in Table 4-1.

Figure 4-1 follows
Table 4-1 follows

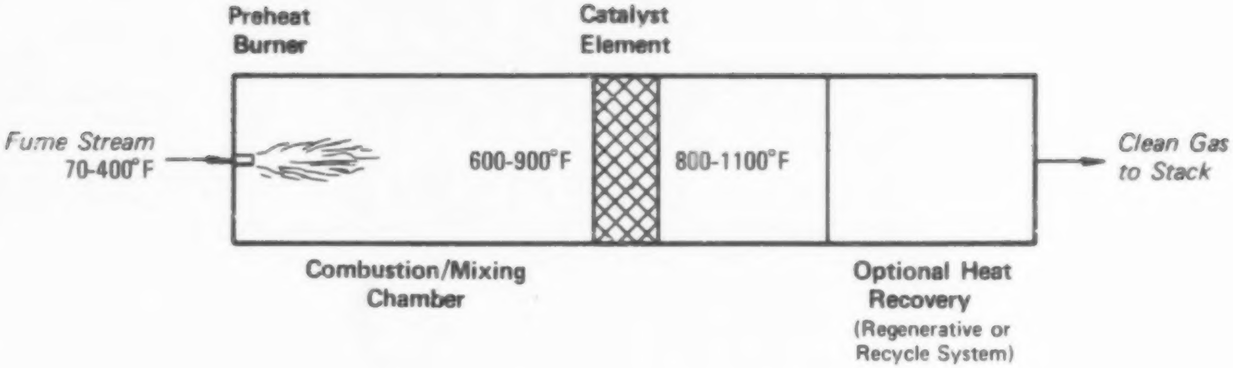

Preheat Burner
Catalyst Element
Fume Stream 70-400°F
600-900°F
800-1100°F
Clean Gas to Stack
Combustion/Mixing Chamber
Optional Heat Recovery (Regenerative or Recycle System)

Figure 4-1. SCHEMATIC OF CATALYTIC AFTERBURNER SYSTEM

56 h

Table 4-1.

Comparison of Thermal and Catalytic Incinerators

Assume:

(1) Process gas volume is 6000 SCFM at 300°F.
(2) Standard solvents for the metal decorating industry: toluene, MEK, xylene, etc.
(3) Design for 90 per cent conversion of hydrocarbon.
(4) Fuel cost is $0.75 per MMBTU.

Item	Thermal	Catalytic
Inlet volume	6000 SCFM 8600 ACFM	6000 SCFM 8600 ACFM
Solvent at 25 per cent LEL	36 gph	38 gph
Operating Temp. F	1350	900
Preheat at zero solvent MMBTU/hr	7.3	4.2
Preheat at full solvent MMBTU/hr	3.0	zero
Inside cross sectional area, sq ft	11.1	6.0
Refractory thickness, in.	9"	5"
Approximate weight, lbs	18,600	12,000
Cost of capital equipment	$18,400	$19,000
Fuel saving		
MMBTU/hr	—	3.1
$/hr	—	2.33
$/5000 hrs	—	11,650.
Catalyst replacement	—	$7750
Payout of catalyst bed	—	3300 hours
Installation ratio	1/1	.75/1
Shipping cost ratio	1	.67/1
Maintenance factor	refractory stack burner	refractory catalyst burner

Table 1 Comparison of temperatures required to convert combustibles to CO_2 and H_2O.

	Ignition Temp. °F			
Combustible	Thermal	Catalytic	Difference °F	Catalyst/Thermal %
Benzene	1076	575	501	53.4
Toluene	1026	575	451	56.0
Xylene	925	575	350	62.2
Ethanol	738	575	163	77.9
MIBK	858	660	198	76.9
MED	960	660	300	68.8
Methane	1170	932	238	79.9
Carbon Monoxide	1128	500	628	44.3
Hydrogen	1065	250	815	23.5
Propane	898	500	398	55.7

WITH PERMISSION FROM: *POLLUTION ENGINEERING*

4.1.3 Limitations

The oxidation performance (conversion of hydrocarbons) of a catalytic system depends on a complex interaction of several factors: catalyst (type, amount, and activity), operating temperature, combustible contaminant type and concentration, and flow velocity of waste stream through the catalyst bed. The amount of catalyst required in a catalytic afterburner depends strongly on the fractional reduction of pollutant concentration which the afterburner must accomplish, rather than on the actual outlet concentration of pollutant. This dependence is such that conversions up to 90 - 95% can be attained with reasonable catalyst volumes. However, the catalyst volume required for very high conversion (e.g., >98%) generally makes catalytic afterburning uneconomic when such performance is required. This limitation can be important in specialized odor control applications where concentrations of strongly odiferous species (e.g., aldehydes or acrylates) must be reduced by factors of 100 or more.

For carbon monoxide and most gaseous hydrocarbons encountered in waste streams, the available catalysts are highly effective in promoting oxidation at temperatures of 600 - 900°F. For methane, however, the catalysts are much less active, and temperatures 300 - 400°F higher are required for substantial conversion. In applications where the waste stream contains methane, and control regulations require significant reduction in methane content to meet overall performance requirements, a catalytic afterburner might have to be operated with an outlet temperature of 1100 - 1200°F to meet these requirements. Under these conditions, the fuel economy advantage of catalytic over thermal afterburners is seriously decreased.

4.1.3.1 Particulate Containing Streams

Oxidation catalysts used in catalytic afterburners are relatively ineffective in treating waste streams containing combustible particulates, unless the particulate material can be totally vaporized in the preheat/mixing region before reaching the catalyst bed. Catalytic afterburning is not recommended for streams containing significant loadings of particulates which cannot be vaporized (either organic or inorganic). The particulates tend to deposit in the catalyst matrix, plugging flow channels, and forming a surface coating on catalytic surfaces which inhibits contact between the waste stream and the catalyst. Catalytic systems have been used successfully for controlling particulates in waste streams from phthalic anhydride plants. In this application, the combustible material (principally phthalic and maleic anhydrides) can be adequately vaporized at reasonable temperatures and residence times in the preheat/mixing section of the afterburner. Although catalytic afterburners performed adequately in this application, most phthalic anhydride manufacturers now use thermal afterburners with heat recovery. Displacement of catalytic systems from this application has been principally on economic grounds. In controlling emissions from asphalt oxidizers, where part of the waste stream hydrocarbon content is an aerosol of high boiling unsaturated oils, catalytic afterburners have been unsuccessful. A part of this failure is due to inability to totally vaporize this oil mist at reasonable temperature and residence time in the preheat/mixing section. The unvaporized oil droplets resulted in a coating of tar or coke forming on the catalyst surface. A contributing factor to failure of catalytic units in this application, however, has been very high and poorly-controlled hydrocarbon contents in the waste stream entering the afterburner, resulting in repetitive exposure of the catalyst to temperatures of 1400°F or higher. Under these conditions catalysts rapidly lost activity. This same factor has also resulted in materials failure in thermal afterburners in this application.

4.1.3.2 Control of Waste Generating Process

It has been indicated above that the rate of decline in catalyst activity and hence the frequency of catalyst replacement, are strongly affected by extreme temperature excursions due to fluctuations in heating value of the waste stream and by particulates and specific catalyst poisons in the waste stream. These factors impose constraints on the waste stream, which may require some accomodation of the process generating the waste stream in the way of control, surveillance or even modification if catalytic afterburning is to be used successfully. Commonly described examples of such process accommodations are: 1) galvanized metal should not be used in process ovens or ductwork between process and afterburner since zinc is a catalyst poison (stainless, aluminized steel or even carbon steel are usually satisfactory); 2) mercury in glass thermometers should not be used in locations where breakage would contaminate the waste stream with mercury vapor; 3) phosphorus from phosphate metal-cleaning detergents and other sources should be excluded from the process; 4) lubricants for oven conveying equipment in solvent evaporation, paint or core baking ovens should be selected so as not to contain silicones or inorganic metal compounds which would contribute to catalyst poisoning on evaporation or degradation. If such constraints are inherently met by the waste-generating process or can be met economically and with a tolerable level of change in operating and maintenance procedures for the process, then catalytic afterburning should be considered on an economic basis as a technically viable control technique.

4.1.3.3 Performance Monitoring

An additional limitation has been frequently raised by the staff of the Los Angeles County Air Pollution Control District (LACAPCD) in explanation of their reluctance to approve installations of catalytic afterburners. This concerns the difficulty of predicting or monitoring performance of an installed unit. With thermal afterburners it is assumed that performance is strictly determined by operating conditions (waste stream flow rate and composition and combustion chamber temperature). Thus, if satisfactory performance has once been demonstrated under a given set of conditions, reproducing these operating conditions will assure such performance. With catalytic units, performance depends not only on operating conditions, but also on the current activity of the catalyst. The air pollution control enforcement agency, and presumably also the afterburner user, would like a continuous or at least frequent method of determining whether performance of a catalytic system is satisfactory. The method should not be more complex or difficult than the measurement of the operating temperature in a thermal unit. For afterburner systems operated to meet odor or nuisance regulations, subjective evaluation of the stack effluent is usually satisfactory. Difficulties arise primarily in applications where performance must meet solvent control regulations such as the LACAPCD Rule 66. If the hydrocarbon concentration in the waste stream is relatively constant, and sufficiently high (>2% of the LEL), the temperature rise of the waste stream between inlet and outlet of the catalyst bed can be used as a measure of the amount of oxidation taking place in the bed. At 2% of the LEL, total conversion would correspond to a temperature rise of 50 -60°F. However, in a catalytic afterburner system, not all of the oxidation occurs in the catalyst bed. As indicated in Figure 4-2, a significant part of the hydrocarbon in the waste stream (10% to perhaps 40 or 50%) is oxidized in passing through the preheat/mixing section. The fraction consumed there is dependent on the preheat temperature and on the chemical nature and thermal oxidation characteristics of the hydrocarbons in the particular waste stream, the flow rate of waste stream, and on the design of the preheat burner and mixing section. Thus, since only a part of the hydrocarbon reaches the catalyst bed still unoxidized, the temperature rise across the catalyst bed can at most correspond to oxidation of this remaining fraction, and will indicate a conversion less than that attained by the overall system. The overall fractional conversion of the system, X_o, may be represented as:

$$X_o = X_p + f_c (1 - X_p) \qquad (4\text{-}1)$$

in terms of X_p, the fractional conversion in the preheat/mixing section, and f_c the fraction of hydrocarbon reaching the catalyst which is converted in the catalyst bed. The final term is directly related to the observed temperature rise across the catalyst:

$$f_c (1 - X_p) = \frac{C_p \Delta T}{Q C_o} \qquad (4\text{-}2)$$

where: C_p is the average specific heat of the waste stream flowing through the catalyst bed (Btu/lb-°F).

ΔT, the observed temperature rise (°F).

Q, the heat of combustion of the hydrocarbon in the waste stream (Btu/lb).

C_o, the initial concentration of hydrocarbon in the waste stream (lb HC/lb waste stream).

If a relation, such as the lower curve in Figure 4-2, giving the fractional conversion in the preheat section, can be supplied by the system manufacturer on the basis of prior experience with similar installations and waste streams; or derived by the user from results of a limited program of analytical tests, then the performance factors X_o and f_c can be obtained from measurements of ΔT across the catalyst bed. Both factors are of importance in determining whether the system is meeting performance requirements, and if not, whether performance can best be improved by raising preheat temperature or by cleaning or replacing catalyst.

The relationship between performance and temperature rise across the catalyst bed is further complicated by the fact that oxidation reactions within the two parts of the system may not be complete (to CO_2 and water as products), but may instead yield significant amounts of carbon monoxide and other partial oxidation products. The heat liberated by the oxidation reaction is then not proportional to the extent of removal of the original hydrocarbon, but depends on the products formed from it as well. The oxidation reactions which take place within the catalyst section under normal operating conditions are invariably complete (to CO_2 and H_2O as products) provided the waste stream has sufficient oxygen content. In the preheat section, the oxidation of hydrocarbon takes place in a temperature range (600 - 1100°F), where thermal processes are known to lead to partial oxidation products. (This is precisely the temperature range used in stack burners for plume abatement where solution of the plume problem frequently creates a worse odor problem due to formation of partial oxidation products.) These often include strongly odorous aldehydes as well as CO. The catalyst then must complete the oxidation of these intermediates as well as oxidize the remainder of the original hydrocarbon. Thus, though at first glance, the curves in Figure 4-2 indicate that the preheat section is contributing significantly to overall system performance in the 300 - 400°C temperature range, in fact the increasing amount of partial oxidation occurring there may be creating a more difficult problem for the catalyst. If the concentration of aldehydes reaching the catalyst is so high as to require 99+% conversion (for odor abatement), a catalyst bed designed for 90 - 95% conversion will not be able to do the job. The most satisfactory solution to this problem is to reduce the partial oxidation in the preheater, not to attempt to get better performance from the catalyst. Since the amount of partial oxidation occurring in the preheater depends on temperature (as well as residence time and burner design), that section should be designed for and operated at the minimum temperature compatible with satisfactory catalyst performance.

From the standpoint of developing a performance monitoring system, what is needed is a curve similar to that shown for the preheater in Figure 4-2 indicating the fraction of the total heat of combustion of the hydrocarbon in the waste stream which is liberated in the preheat section (rather than the fractional destruction of the original hydrocarbon). Development of such a curve would have to be based on experience either by the system user or the manufacturer. With such a curve, the equations presented above could be used to characterize performance if X_o and X_p were interpreted as fraction of total available heat of combustion rather than fractional conversion of the original hydrocarbon in the waste stream. Such a performance monitoring system appears complicated to initiate. Once set up it should be fairly simple to use. However, such a system is valuable only when hydrocarbon content of the waste stream (Co) is known and can be held relatively constant. A much more generally applicable monitoring system would require development of an inexpensive analytical instrument which could be used either in-line or intermittently for determination of total hydrocarbons and CO in the waste stream and in the afterburner effluent. Such an instrument would be of particular value in overcoming problems of performance monitoring for catalytic systems, but would also find application in thermal afterburners used under a wide range of waste stream concentrations and compositions.

Figure 4-2 follows

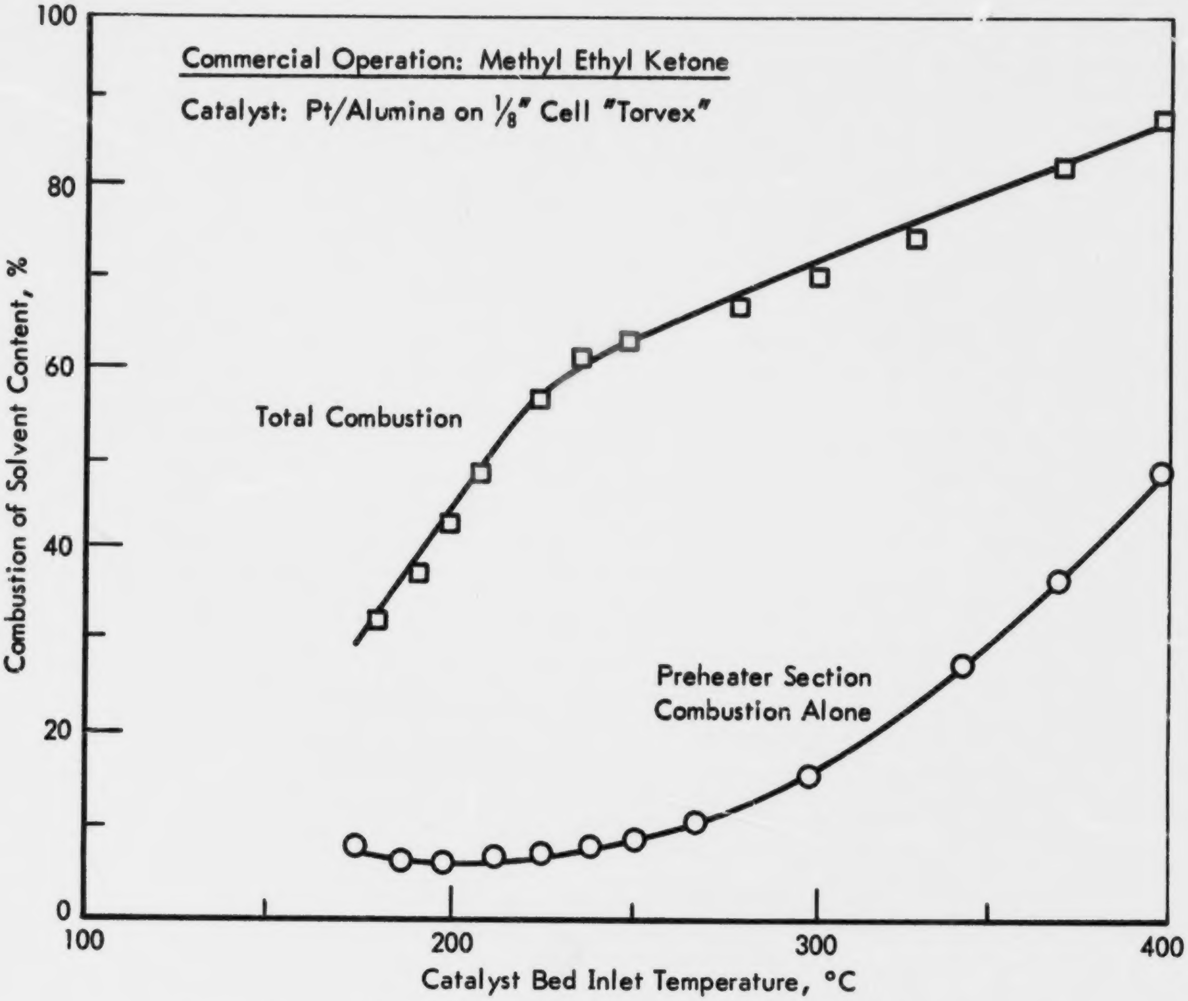


Figure 4-2. CONTRIBUTION OF PREHEAT SECTION TO OVERALL CONVERSION IN A CATALYTIC AFTERBURNER
(From Fig. 4-8)

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4.1.4 Applications

Manufacturers' brochures and trade journal articles have frequently quoted extensive lists of applications for which catalytic afterburners are recommended. For some broad categories of applications, considerable development work has been done by manufacturers and users of catalytic systems. The probability of success in using catalytic afterburning can be reasonably predicted within these categories--the problems and limitations inherent in the process generating the waste stream are reasonably well understood. For new applications to totally different processes, it is considerably more difficult to predict whether operation will be successful. Most equipment manufacturers maintain at least laboratory facilities and several have portable pilot units for on-site testing in unusual or new applications. Such tests are useful particularly in determining temperature and catalyst volume required with fresh catalyst, but seldom can a sufficiently extensive program be carried out to determine whether catalyst life of more than a few months can be expected. The demonstration of the capability of catalytic afterburning in a new application usually requires the installation of a full-size (though perhaps small) unit in the field. Such an installation requires assumption of considerable risk by the user (which may be shared by the equipment manufacturer). However, if the unit proves successful in the application, its operating costs can be considerably lower than a corresponding thermal unit.

Until a few years ago, most applications of catalytic afterburners were to streams with relatively high hydrocarbon contents. Metal decorating (lithography and coating ovens) and paint drying ovens have been subjects of extensive field trials of catalytic afterburners. Applications have often been successful, (though less often when destruction of methane by the afterburner is required),where equipment manufacturers are knowledgeable concerning the process factors which have in some instances lead to short catalyst lives in these applications. In wire enameling ovens, also considerable field experience has been acquired. Here, due to high temperature operation of the enameling oven, and the volatilization of metal contaminants in the process, very short catalyst lives have been experienced with noble metal catalysts. Applications work is continuing in this area with emphasis on possible use of catalysts less subject to deactivation by phosphorus and copper poisoning. Catalytic afterburners have been used fairly extensively and successfully in controlling emissions from varnish cookers, and to some extent in foundry core baking. It is generally agreed by both equipment manufacturers and users that fumes from asphalt blowing operations (and probably also from asphalt impregnation processes) are not an appropriate application for catalytic units.

More recently, engineering and development work by equipment manufacturers has centered on new applications involving relatively low concentrations of hydrocarbons in the waste stream to be afterburned. In these cases, the afterburner is usually installed to control odor or other nuisance problems. Several successful installations have been made on the effluent air streams from dryers following web-offset lithograph presses in the printing industry. Effluents from various food processing operations have also been sucessfully treated by catalytic afterburning. However, the particulates associated with effluents from coffee-roasting and some other food drying operations would be expected to make these unsuitable for catalytic afterburning.

Some large installations in the chemical industry have been quite successful, technically at least. Catalytic systems are used as emission controls on off-gas from ethylene oxide plants and have been used on off-gas from phthalic anhydride plants. Probable success in applications to waste streams from chemical plants is usually somewhat easier to assess than in other applications. Such streams are well characterized in terms of chemical composition, and the presence of contaminants which might decrease catalyst life is usually known in advance of deciding to install the unit. The extensive use of catalytic systems for abatement of nitrogen oxides in tail gas from nitric acid plants is outside the scope of this study. However, such systems are similar in design and use similar catalysts, and are designed and supplied by the same manufacturers as are oxidative catalytic afterburners. In some of these chemical process applications, the waste stream is available under pressure. In such cases, the system must be designed to accommodate this condition. In the equipment descriptions which follow, the more normal application in which the waste stream is essentially at ambient (atmospheric) pressure is assumed.

The preceding discussion of applications is not intended to be exhaustive, but rather to point out general areas of applications where considerable experience has been acquired. In considering installation of a catalytic afterburner, the customer should discuss in detail with equipment manufacturers, their experiences with systems in applications most similar to his own.

4.2 Performance Characteristics

In the preceding section, the contribution of oxidation in the preheat/mixing section to the overall performance of a catalytic afterburner was discussed in some detail. Recognizing that some significant fraction of the total hydrocarbon in the waste stream may have been oxidized in that upstream section, here the discussion will be restricted to the factors determining the extent of oxidation in passing through the catalyst bed itself. The performance characteristics of catalysts from different manufacturers are different. Some attempt will be made to point out major differences both in this discussion and in the section on equipment description. However, the emphasis in the present discussion is on characteristics common to all of the catalysts available for afterburners.

Performance requirements for afterburners may be stated either: 1) in terms of a maximum outlet concentration of contaminant in the cleaned waste stream; or 2) in terms of fractional conversion of the contaminant from the inlet waste stream. Nuisance and odor restrictions are most easily stated in terms of maximum outlet concentrations, while solvent control regulations are usually stated as required fractional conversion. The two methods of stating performance are simply related provided the inlet contaminant concentration is known. In dealing with performance of catalytic systems, the fractional conversion expression is more naturally related to the design and operating variables of the system. In most applications, requirements demand conversions in the catalyst bed of at least 70 - 80% (to give overall system conversion of 85% or higher). Fractional conversions higher than about 95% in a catalytic bed require uneconomically large volumes of catalyst. Thus, the range of conversions of interest is perhaps 70% to 95%.

4.2.1 Rate of Oxidation

For all of the catalysts in commercial use, the rate of oxidation of a combustible component of the waste stream at any point within the catalyst bed is well described as being first order in (proportional to) the concentration of the component in the waste stream at that point. As a consequence, the overall conversion of that component can be expressed as:

$$-\ln(1 - f_c) = \frac{\bar{k}_{eff} \cdot v_{cat}}{V_{gas}}$$

where:		
	ln	natural logarithm
	f_c	fractional conversion of component
	$\bar{k}_{eff}$	average effective first order rate constant for oxidation of the component (sec^{-1})
	v_{cat}	volume of catalyst in bed (ft^3)
	V_{gas}	volumetric flow rate of waste stream (actual ft^3/sec.)

The rate of oxidation and conversion are essentially independent of the oxygen concentration in the waste stream, provided this stream after preheating contains at least 2%v oxygen in excess of that required for complete oxidation of all combustibles present. The conversion can then be thought of as determined by the product of two factors: 1) the amount of catalyst per unit volume of gas being treated per unit time; and 2) an effective rate constant for the oxidation process. The amount of catalyst is selected in the design of the unit, and hence is not an operating variable. However, its influence on performance is important in determining the operating capability of the system. Figure 4-3 shows the effect of the (relative) amount of catalyst used on conversion performance, when the comparison is made for constant effective rate constant (constant operating conditions and catalyst activity). The figure shows that twice the amount of catalyst is required for 90% conversion as for 66%. A further increase of catalyst amount by 50% will then raise the conversion over 95%.

The effective rate constant for oxidation is directly controlled by operating conditions. Figure 4-4 shows the typical effect of temperature on conversion through its effect on the rate constant for oxidation. The overall oxidation process is comprised for a complex sequence of steps. However, for examining the effect of operating variables (and indeed even for correlation and prediction of performance), these may be lumped into two process which appear to dominate the determination of the oxidation rate. The first process is a chemical reaction taking place heterogeneously at the surface of the catalyst element, and having the usual strong dependence on temperature characteristic of chemical reaction rates. It is recognized that, at least with the Pt/Al_2O_3 catalysts, the reaction involved occurs at Pt crystallite sites distributed within the porous alumina coating on the matrix. Thus, its rate involves diffusion of reactants and products through this porous layer as well as the actual chemical reaction rate. However, in the normal range of operating condition, it appears possible and reasonable to treat this complex process as a chemical reaction rate, proportional to the geometric surface area of the matrix, and having a constant activation energy (temperature dependence coefficient for the chemical rate constant). The second process is mass transfer of the species to be oxidized (hydrocarbon) to the surface of the catalyst from the bulk gas stream; and the same process of mass transfer of oxidation products (CO_2, H_2O) from the surface back into the bulk gas stream. Typical of such mass transfer processes, the rate of this process is relatively slightly affected by temperature. However, with at least some of the catalyst configurations used it is affected by other design and operating conditions, in particular the velocity of gas flow through the matrix. These two processes operate in series. At low temperature, where the chemical rate process is far slower than mass transfer, the overall rate is determined by the chemical process, as indicated by the region A-B in Figure 4-4. At high temperatures in region C-D, the chemical rate has become much higher than the mass transfer rate and the overall process becomes limited or controlled by the mass transfer rate. In the intermediate region, B-D, both processes contribute to determination of the overall rate. Figure 4-5 shows the rate constants for these two contributing processes, and their combination into an overall effective rate constant for the oxidation process. In Figures 4-4 and 4-5 the numerical values of conversion, rate constants and temperature were selected for a catalyst supported on a ceramic honeycomb matrix under typical operating and design conditions. For other catalyst types and design conditions the same qualitative behavior with operating temperatures will be observed. However, the temperature region and the conversion level at which transition between chemical and mass transfer control takes place

Figures 4-3, 4-4, and 4-5 follow

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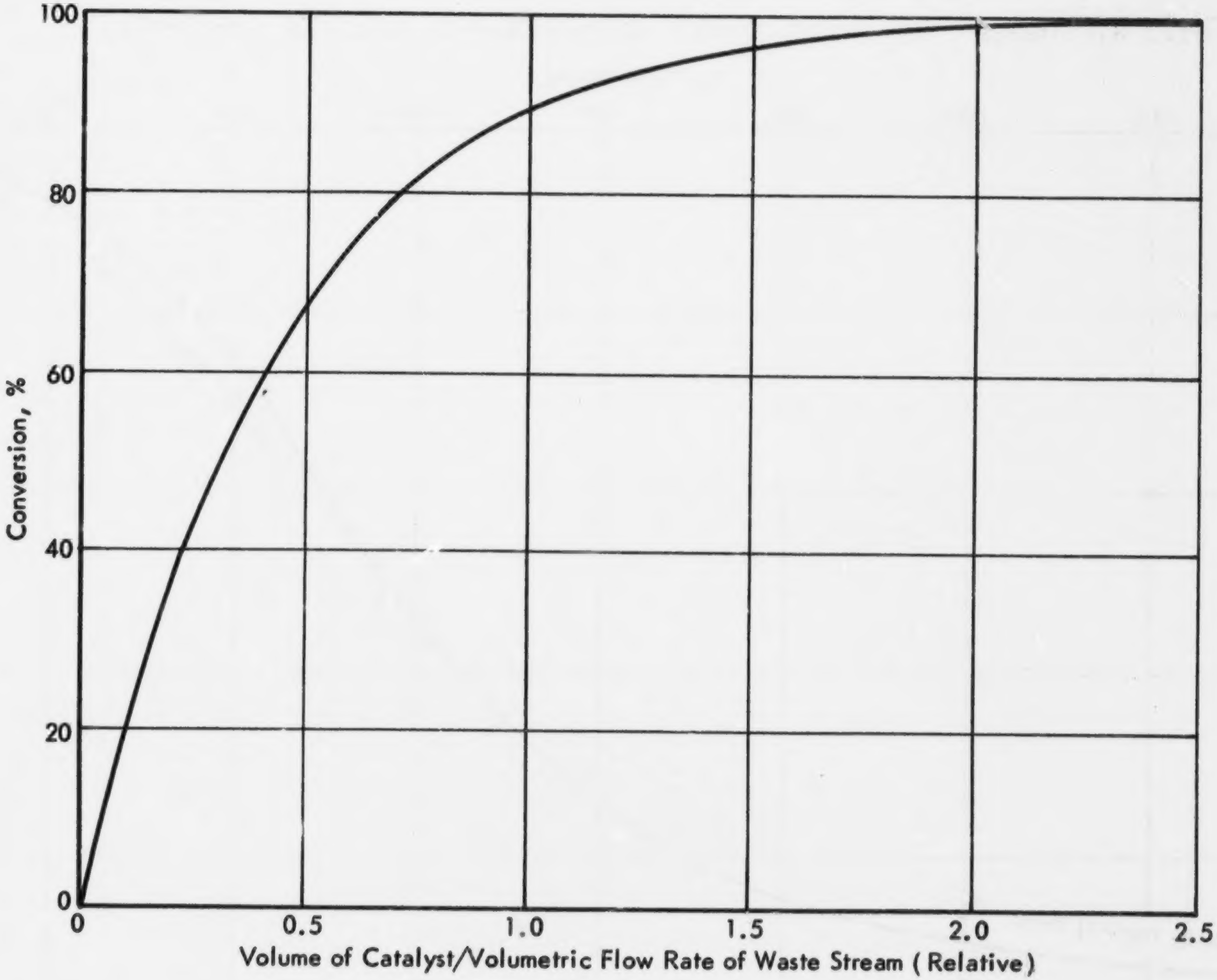

Figure 4-3.

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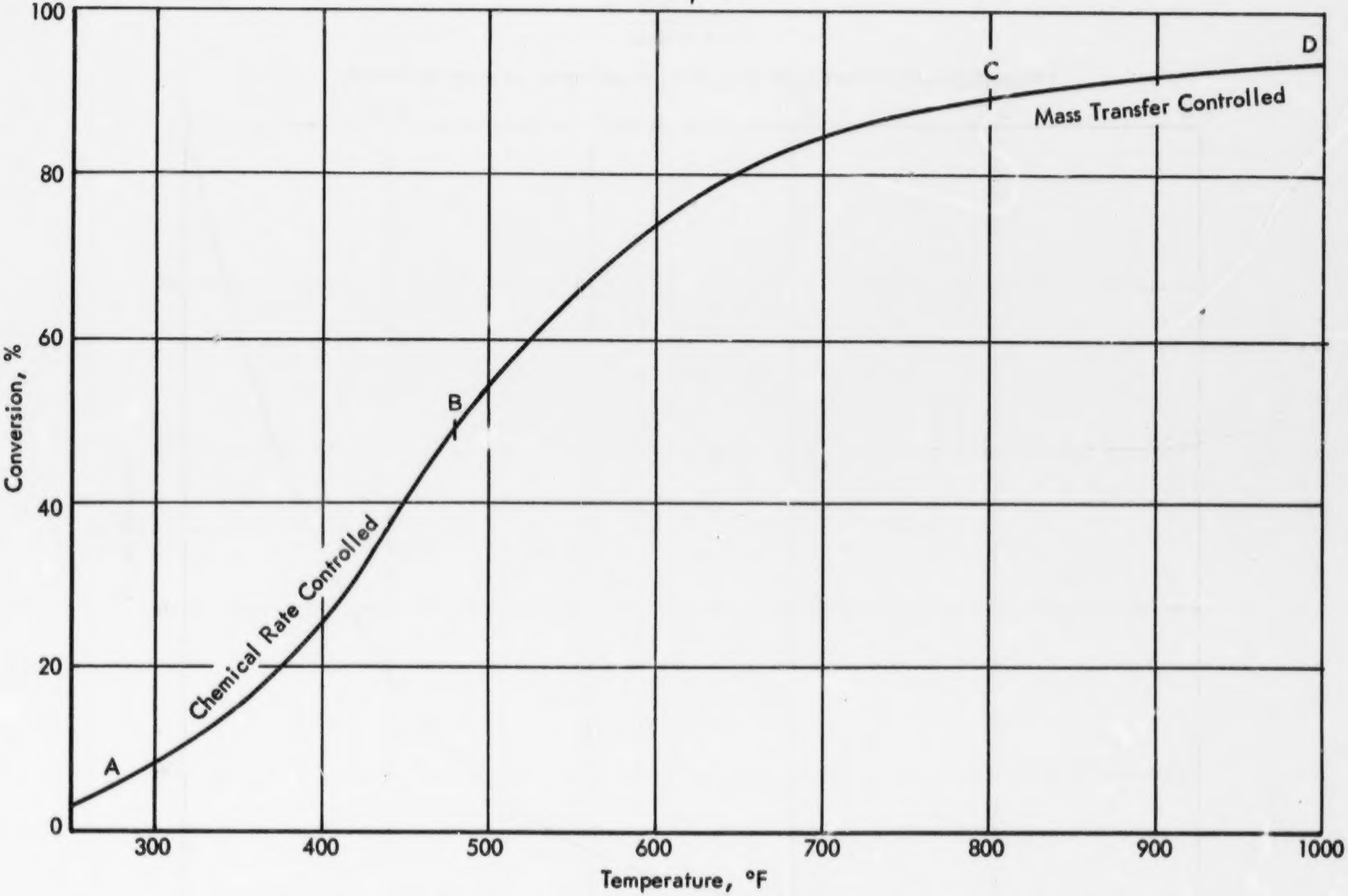

Figure 4-4. EFFECT OF TEMPERATURE ON CONVERSION OVER AFTERBURNER CATALYST

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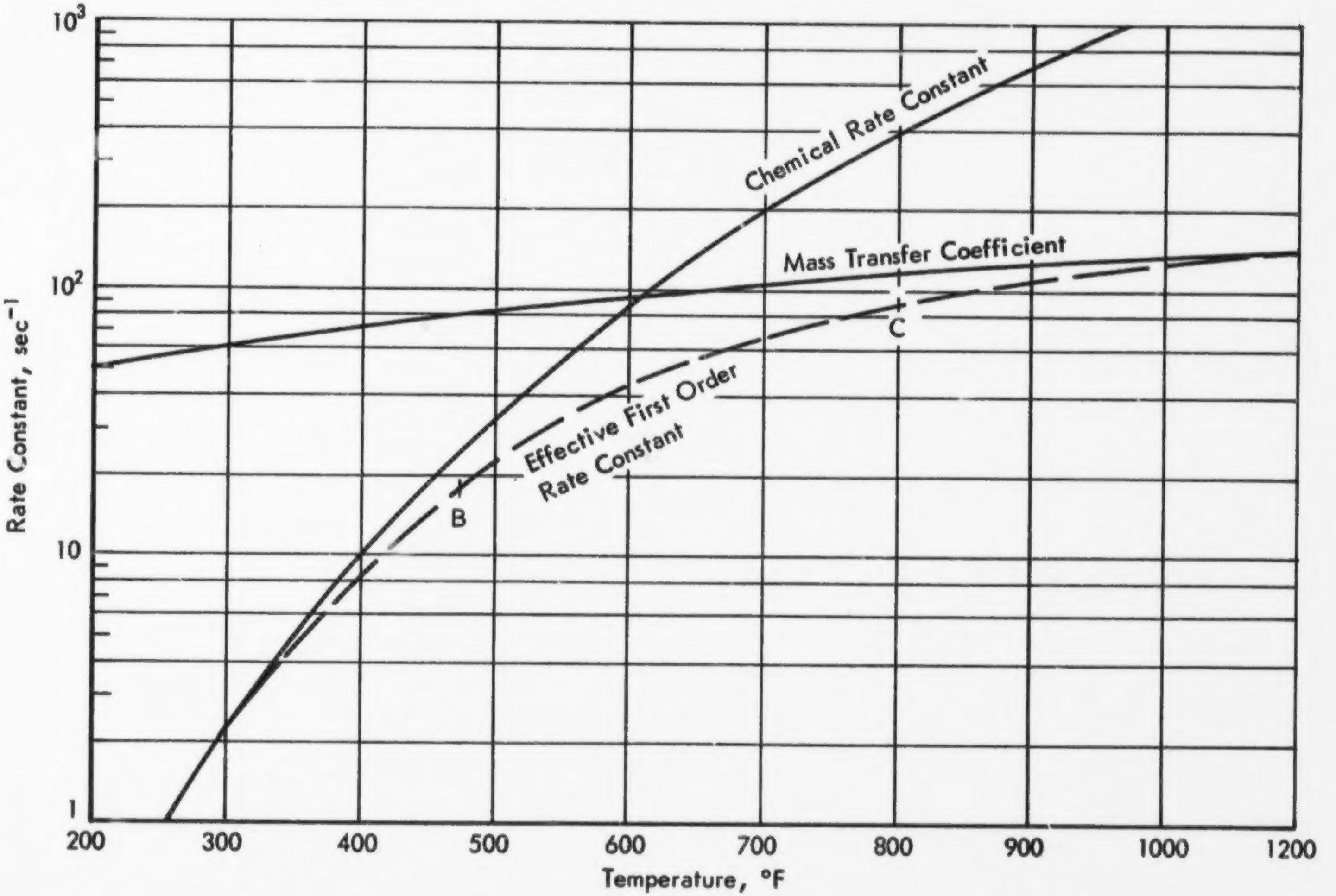

Figure 4-5. EFFECTIVE FIRST ORDER RATE CONSTANT AND CHEMICAL AND MASS TRANSFER RATES CONTRIBUTING TO IT

will be shifted. The design operating point (i.e., quantity of catalyst and operating temperature) is frequently selected to meet required performance at or near point C on a curve similar to Figure 4-4. Such an operating point makes maximum use of preheat temperature in increasing conversion and minimizes the amount of catalyst required. At such a point, the system performance will be only slightly sensitive to changes in temperature. In particular, a significant improvement in performance over the original design cannot be obtained without a rather large increase in temperature. For the design illustrated by the figure, a 200°F increase from point C (800°F) to point D (1000°F) raises conversion from 89% only to 93%.

If the catalyst is not sufficiently active for oxidation of the particular component desired (e.g., methane), then sufficient catalyst must be used to attain the required conversion that the mass transfer controlled region is shifted to an impractically high temperature. In such a case the design operating point would be nearer to or in the region of chemical rate control and hence performance at the design point would be more sensitive to temperature. However, significant improvement in performance (over design) by raising temperature might still be constrained by the need to keep catalyst and other system components below their maximum permissible temperatures. The overall conclusion concerning the effects of operating temperature are that it provides a moderate leverage for affecting conversion performance, but that a major improvement (over design) in the performance of a catalytic system could not normally be obtained by raising temperature, but would require an increase in the amount of catalyst used in the unit.

4.2.2 Effect of Combustible Molecular Type

The catalysts used in afterburners are effective in promoting oxidation for a wide variety of gaseous, combustible materials. However, their specific activity does depend on the type of molecule being oxidized. Figure 4-6 shows conversion-temperature curves for H_2, CO, CH_4 and solvent hydrocarbons for a typically designed system using Pt/Al_2O_3 supported on ceramic matrix. The catalyst is most active for H_2 oxidation, with significant oxidation below 100°F and high conversions attainable at 300 - 500°F. CO shows significant reaction at 300°F and high conversions at 550 to 800°F. Most of the common solvent hydrocarbons lie within a narrow band with commercial alumina-supported, noble-metal catalysts. These show significant activity at about 400 - 450°F, and high conversions require 750 to 1000°F. The catalyst is least active for methane with little activity below 750°F and high conversions requiring temperatures of 1100 - 1300°F. The same relative positions of these curves probably also apply to the all-metal, Pt or Pt-Pd supported on nichrome ribbon catalysts. However, these catalysts show a much broader range of activity for solvent hydrocarbon molecules. They have comparable or perhaps even higher activity than the alumina catalysts for a few more easily oxidized species (such as toluene and xylene), but for more difficultly oxidized hydrocarbons (such as benzene and n-heptane) temperatures as much as 400 - 500°F higher would be needed. With these all-metal catalysts, design and operating conditions depend strongly on the particular composition of the waste stream being treated. In general, their design points tend to be nearer to the region of chemical rate control than those for the alumina-base catalysts.

Table 4-2 presents a collection of ignition (catalytic) temperatures and preheat temperatures for oxidation of hydrocarbon and other species over various commercial catalysts. The columns labelled ignition temperature represents the temperature at which 20 - 50% conversion would be attained under normal design conditions. Similarly, the preheat temperatures for 90% conversion refer to a normal design ratio of catalyst volume to waste stream flow rate. However, these preheat temperatures usually refer to oxidation of a stream containing a significant hydrocarbon concentration (10% of LEL), which results in a substantial temperature rise (about 250°F) in passing through the catalyst. Thus, the average temperature of the catalyst bed to attain 90% conversion may be 150 - 200°F higher than the tabulated preheat temperature. These are labelled minimum temperatures because they refer to fresh catalysts. Presumably as catalyst activity declines in use due to aging and poisoning preheat temperature would have to be raised.

An indication of the effect of solvent molecule type on oxidation activity of the all-metal catalysts is given in Figure 4-7. The rate constants, presented on a relative scale in this figure, correspond to the "effective first order rate constants" shown in Figure 4-5. For toluene-xylene, the rates are essentially mass transfer rates, but for the species more difficult to oxidize, the rates shown are more nearly the rates of chemical oxidation.

Figures 4-6 and 4-7 follow

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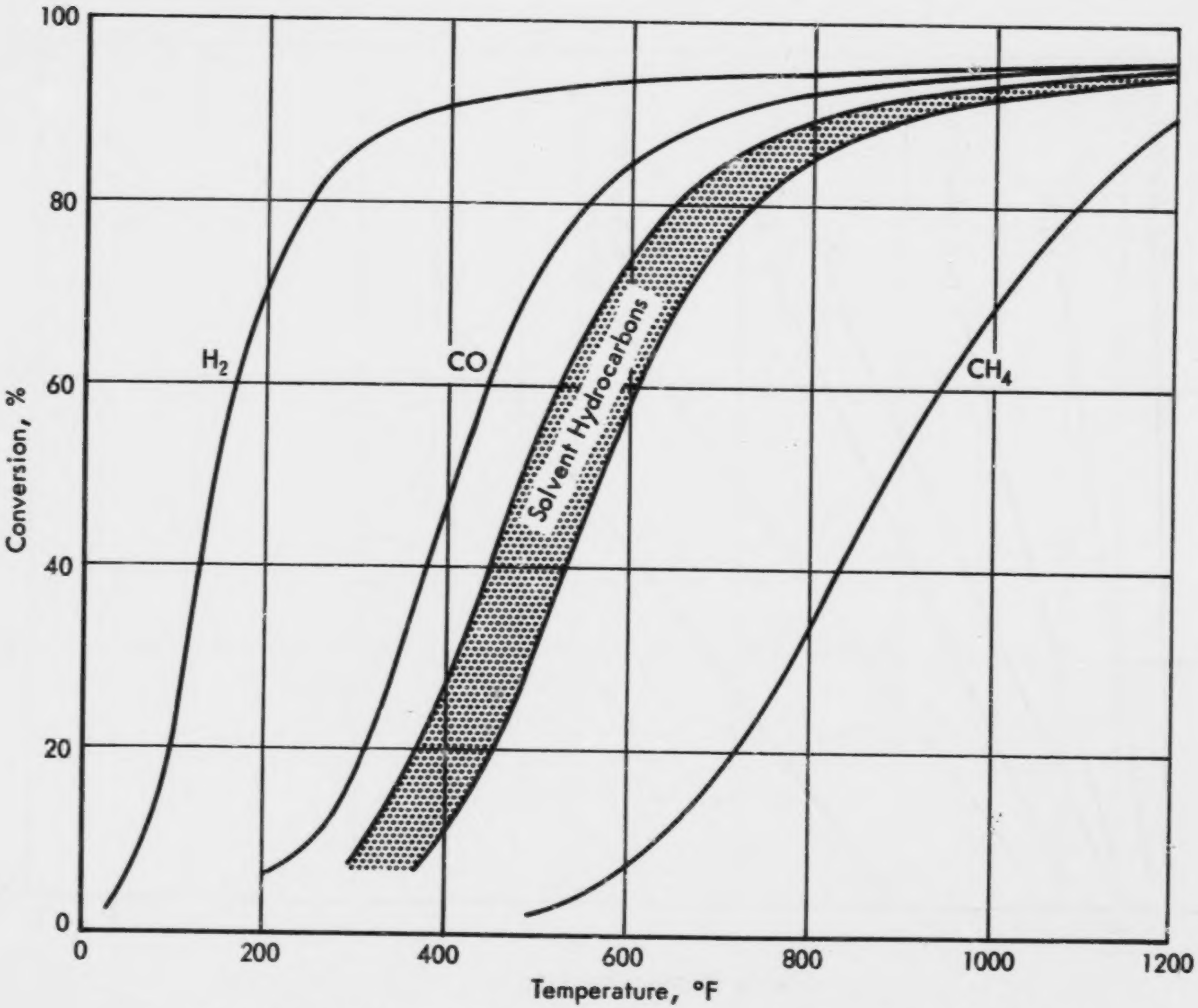

Figure 4-6. TYPICAL TEMPERATURE-PERFORMANCE CURVES FOR VARIOUS MOLECULAR SPECIES BEING OXIDIZED OVER Pt/Al_2O_3 CATALYSTS

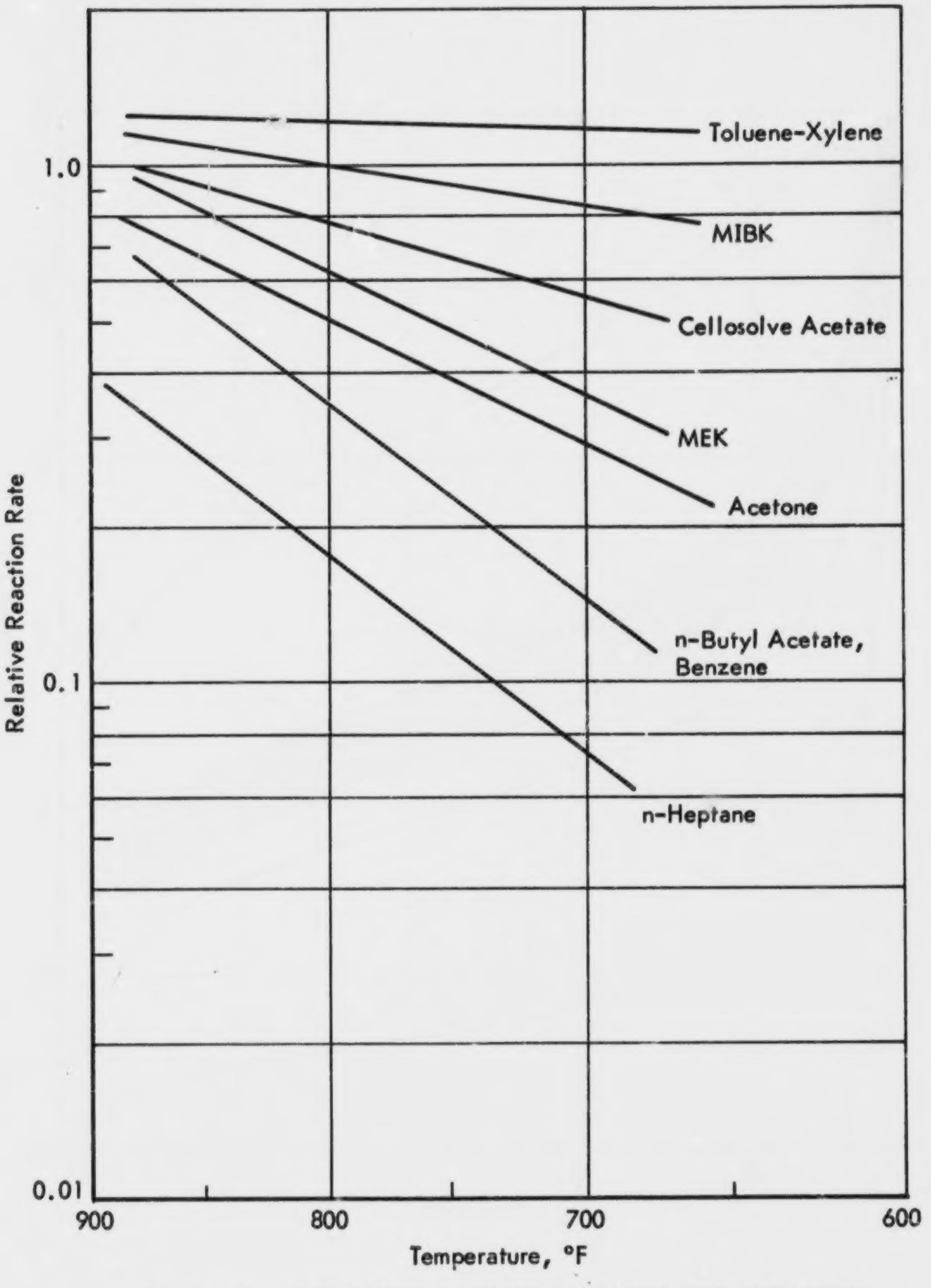

Figure 4-7. CATALYTIC OXIDATION RATES FOR SOLVENTS OVER ALL-METAL CATALYSTS
(Ref. 4-5)

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Table 4-2. TEMPERATURES FOR CATALYTIC OXIDATION

Component	Al_2O_3 Based Catalysts					Nichrome Ribbon Based Catalysts Ignition Temp °F (Suter, 4-12))
	Ignition Temperature °F (Acres, 4-1))	Minimum Preheat Temperature for 90% Conversion with Solvent Concentration 10% LEL				
		(Acres, 4-1))	(Miller & Sowards, 4-8))	(Thomiades, 4-13))	(Romeo & Warsh, 4-10))	
H_2	68	68		250	32	
CO	300	300-390		500	600	
Propane				500	660	650
n-Pentane						590
n-Heptane	320	480-570	480		570a)	580
n-Decane						500
n-Tetradecane						550
Benzene	355	480-570		575	500	440
Toluene	340	480-570		575		460
Xylene	390	480-570	572	575	570	470
CH_4	750	840-930		932	800	760
Ethanol	320	480-570	536	575		
MIBK	345	570-660	544	660		
MEK	345	570-660	544	660		
Mesityl Oxide	355	480-570				
Ethyl Acetate	525	750-840	734			
Dimethyl Formamide	390	660-750				
Pyridine	765	750-840				
Thiophene	635	750-840				
Chlorobutane	800	840-930				

a) Naphtha.

4.2.3 Effect of Hydrocarbon Concentration

In the preceding discussion of operating characteristics, the catalyst operating temperature has been characterized as a single, constant value throughout the catalyst bed. This would be the case if the waste stream contained such a low concentration of hydrocarbon (combustible) as to have a negligible heating value. In many applications (e.g., metal decorating, paint baking) the waste stream may have a significant solvent concentration, up to 25% of the lower explosion limit (LEL). In such a case, as the combustible contaminant is oxidized, the heat of combustion raises the waste stream temperature. The catalyst surface temperature normally follows closely the temperature of the gas flowing over it. The catalyst bed then operates with a temperature gradient in the direction of flow, increasing from the preheat temperature at the inlet end to a temperature as much as 700°F higher at the outlet. For most hydrocarbons and other solvents, the heat of combustion (oxidation) is about 0.5 Btu/SCF for each percent of LEL hydrocarbon concentration. This raises the waste stream temperature about 27.5°F for each percent of the LEL on oxidation. Thus, at 10% LEL concentration into the catalyst bed, the fume stream would have a heating value of 5 Btu/SCF, and the stream temperature would rise 250°F if 90% of the hydrocarbon were oxidized in passing through the bed.

Figure 4-8 shows the effect of solvent concentration on the required preheat temperature for a given conversion level. The lowest curve, labelled "0% LEL" is from the band for "solvent hydrocarbons" in Figure 4-6. It represents preheat temperature required when the entire catalyst bed is isothermal. The upper three curves show the progressive and significant lowering of preheat temperature for a given required conversion. This is brought about by the increase in temperature rise through the bed with increasing hydrocarbon concentration (and heating value) of the waste stream. Thus, the catalyst surface temperature (which determines the chemical rate of oxidation) operates at progressively greater temperature increments above the preheat temperature as the solvent concentration increases. It should be noted that the ultimate limit on performance at high temperatures is still the mass transfer of reactant to the catalyst surface. This process is only slightly affected by the higher temperature resulting from high solvent concentration in the waste stream.

When performance is viewed in terms of catalyst outlet temperature, the temperature-performance curves for various solvent concentrations are brought much closer together. As shown in Figure 4-9, their order with solvent concentration is now reversed, with the highest solvent concentration requiring the highest outlet temperature for a given conversion. With neglible concentration or heating value in the stream, the entire bed operates at the outlet temperature. As the heating value is increased, the temperature at the inlet end of the bed is decreased below the outlet temperature, and is characterized by a lower chemical rate constant. To raise the average temperature and rate up to the level for required performance then means that the outlet temperature must be increased. For concentrations up to 10% LEL (5 Btu/SCF), the curves spread a maximum of 60°F. Outlet temperature in that range is a fairly good indication of performance nearly independent of solvent concentration. At higher solvent concentrations, the spread from the 0% LEL curve becomes quite significant (up to 165°F) and solvent concentration must be considered in setting outlet temperature for desired performance.

Figures 4-8 through 4-10 follow

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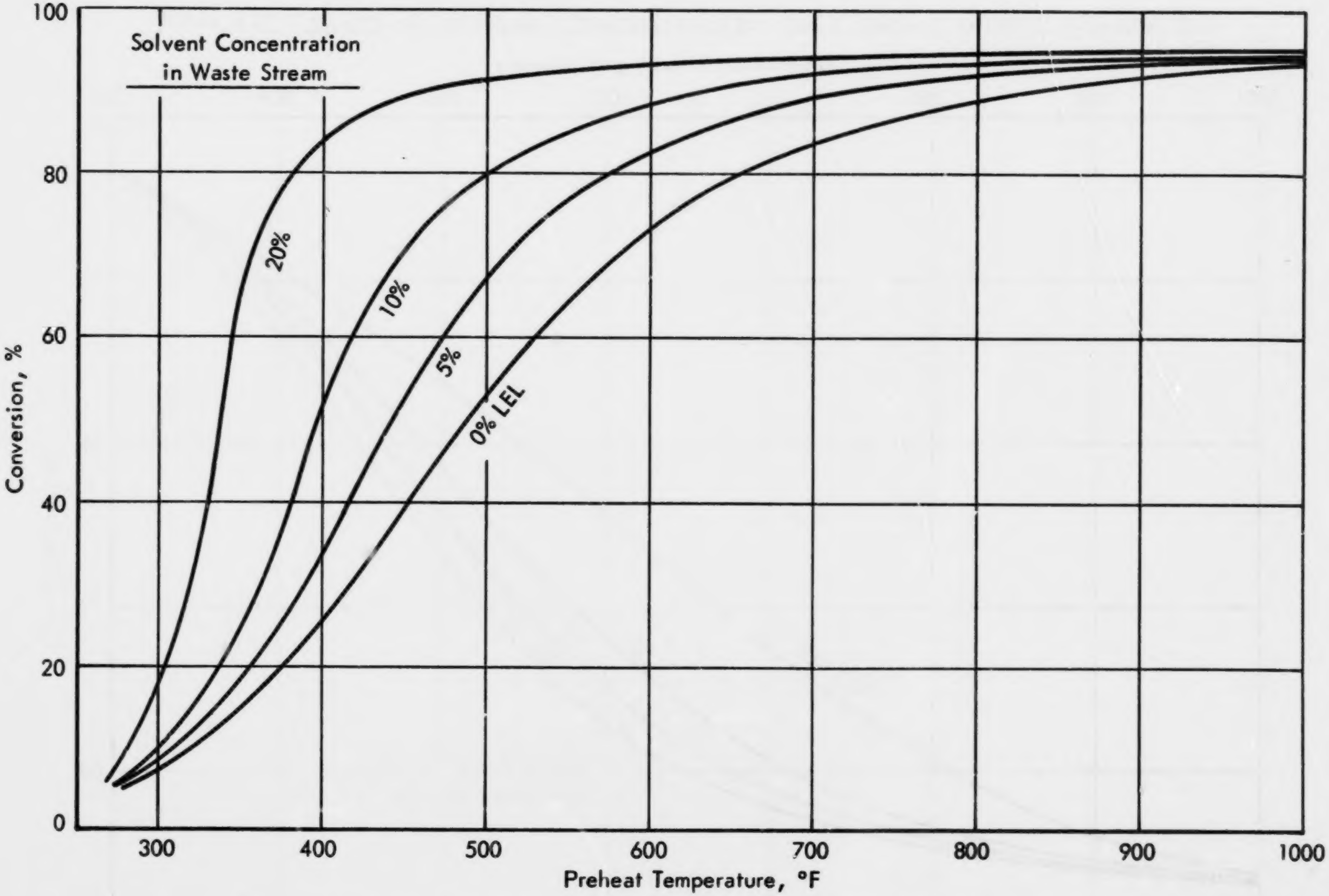

Figure 4-8. EFFECT OF SOLVENT CONCENTRATION ON REQUIRED PREHEAT TEMPERATURE

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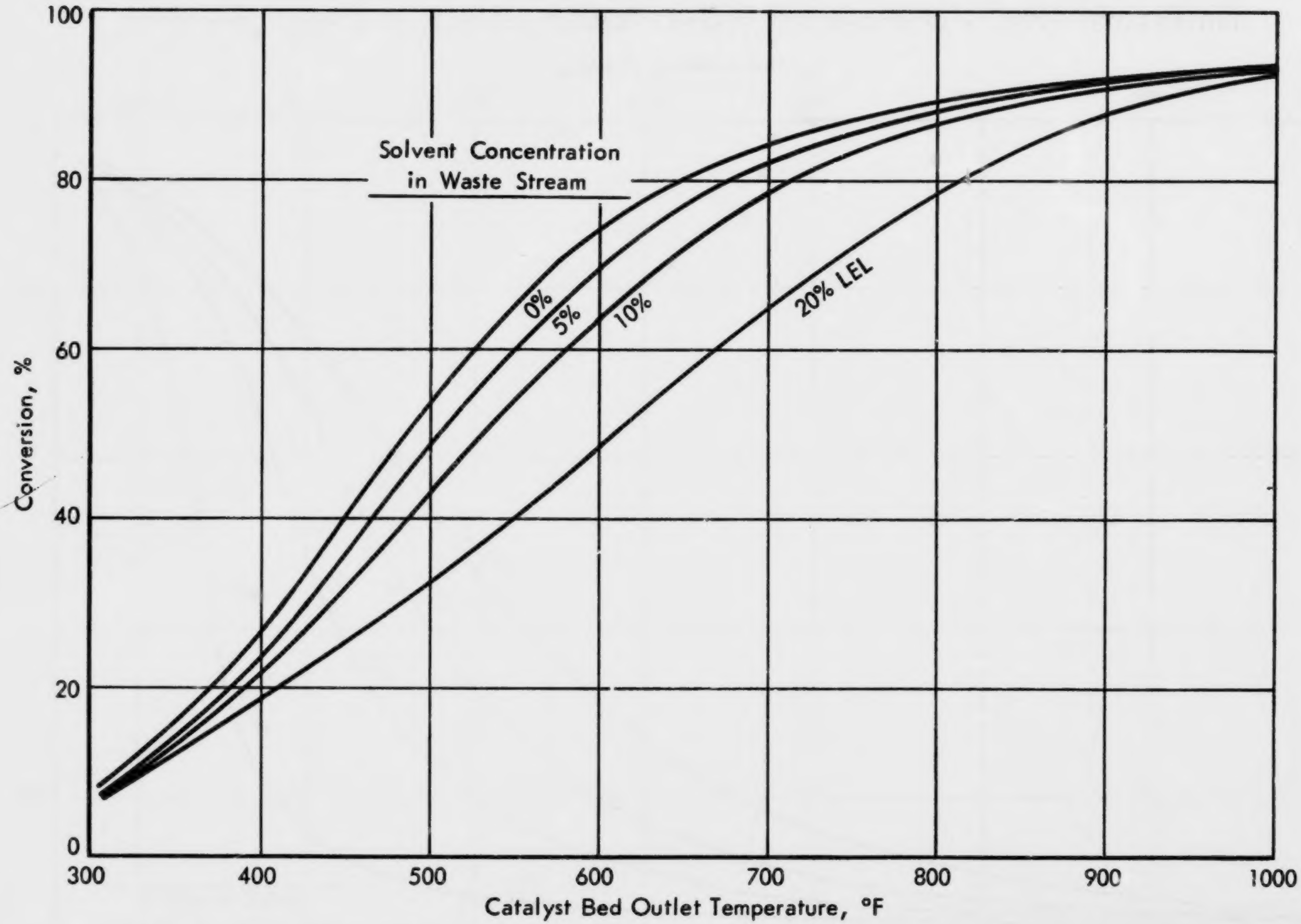

Figure 4-9. EFFECT OF SOLVENT CONCENTRATION ON CATALYST OUTLET TEMPERATURE

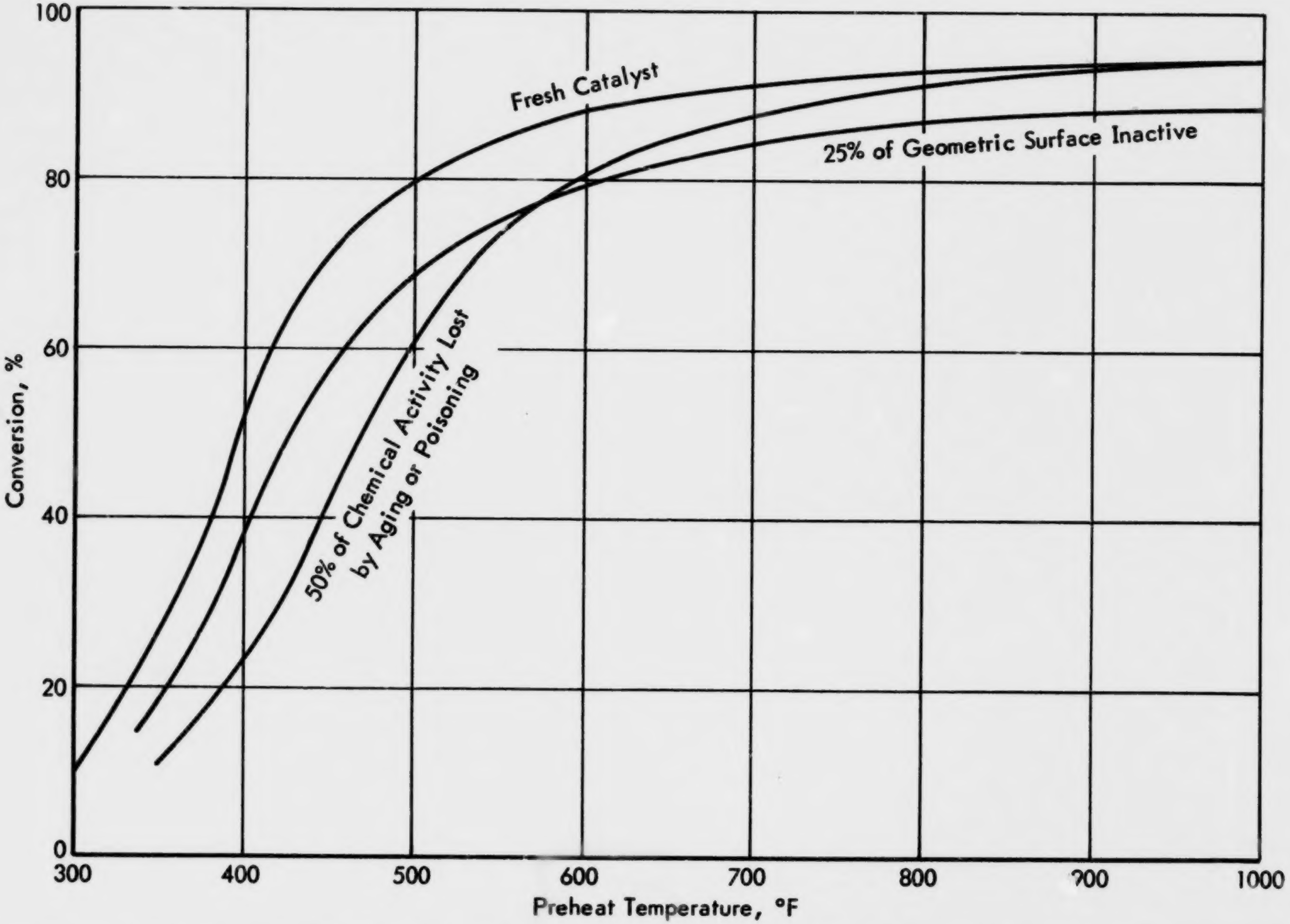

Figure 4-10. PERFORMANCE LOSS ASSOCIATED WITH DEACTIVATION AND POISONING OF CATALYST

Hydrocarbon, 10% LEL on Al_2O_3/Pt Catalyst

4.2.4 Catalyst Deactivation and Poisoning

With time and exposure to the environment within the afterburner, the activity of the catalyst for promoting oxidation will decline. Unless compensating changes in operating condition or maintenance is done, the afterburner performance will decline along with catalyst activity. Compensation for decreased catalyst activity may be made by: 1) initial overdesign in specifying the amount of catalyst needed to attain required performance; 2) raising preheat temperature as chemical activity decreases; 3) cleaning or reactivation of the catalyst; 4) replacement of the catalyst with a fresh charge. The appropriate compensating action depends upon the cause and nature of the deactivation process.

4.2.4.1 Deactivation by Aging or Thermal Processes

In the absence of more specific deactivation processes, catalyst life is limited by thermal aging--micro-structure changes in the active metal crystallites and the porous alumina supporting them--and by loss of active coating from the matrix structure by erosion, attrition and vaporization. With proper operating temperature and good control of temperature, these processes are normally slow and satisfactory performance can be maintained for three to five years before replacement of catalyst is necessary. With both the all-metal and alumina base catalysts, these aging processes are accelerated by increasing temperature. The alumina-base catalysts are somewhat more sensitive in this respect. The three to five-year life estimate is based on operation with a maximum temperature in the catalyst bed (normally at the outlet end) of 1100°F or less for Pt/Al_2O_3 catalysts. If this temperature is raised to 1250 - 1300°F effective life will drop to about one year. Above this temperature, deactivation processes accelerate rapidly, and even short-term exposure to temperatures above 1400 - 1500°F can result in drastic (nearly total) loss of catalyst activity. The manufacturer of the all-metal catalysts specifies 1500°F as the maximum temperature of exposure. Although both types of catalysts are also used in NO_x abatement in tail gas from nitric acid plants at temperatures up to 1400°F, shorter catalyst life is acceptable in that application, and the reducing conditions in units there (as opposed to the oxidative environment in catalytic afterburners as discussed here) are less conducive to thermal aging processes. Regular, preventative maintenance of thermocouples and the control instrument in the temperature control system of catalytic afterburners is recommended since a failure of temperature control could result in deactivation of the catalyst due to overheating. Similarly sufficient control over the process generating the waste stream must be exercised that sudden surges in combustible concentration, which might generate sufficient heat to damage the catalyst are avoided.

The normal consequence of aging or loss of active coating is a loss of chemical activity uniformly over the matrix surface through loss of active sites randomly located throughout the bed. Figure 4-10 compares the performance of fresh catalyst with that for catalyst which has lost half its initial chemical activity by random loss of active sites. The effects of activity loss are quite evident in the low temperature region of chemical rate control. However, they decrease at higher temperatures where the rate is controlled by gas to solid mass transfer. If the initial design point were 90% conversion with a preheat temperature of 650°F, the loss of half the activity of the catalyst would only decrease conversion to

85% at this preheat temperature. Conversion could be restored to 90% bv raising preheat 750°F. This policy of raising preheat temperature to compensate of loss of activity is normally followed until the added fuel cost for preheating makes catalyst replacement a more economically attractive choice, or the outlet temperature has reached a point where catalyst activity is rapidly declining.

A second type of performance change resulting from another pattern of catalyst deactivation is also illustrated in Figure 4-10. In this case, catalyst has been totally deactivated in only one part of the bed. This pattern could result from high temperature exposure of the inlet end due to flame impingement from the preheat burner (poor design of the preheat/mixing section or overloading of the preheat burner), or high temperature at the catalyst bed outlet due to excessive heating value (solvent concentration) in the waste stream. The performance pattern is equivalent to operation with less than the design amount of catalyst. Performance is decreased not only in the low temperature region, but also in the mass transfer controlled region. A total loss of activity in only 25% of the bed would decrease conversion from 90% to 82% at 650°F preheat and restoring performance to 90% conversion would require a preheat temperature in excess of 1000°F. For this pattern of deactivation, raising the preheat temperature may be an impractical policy. In this example raising preheat above 1000°F would raise the outlet temperature (with a hydrocarbon concentration of 10% of LEL) above 1250°F and lead to rapid deactivation of the remaining catalyst.

4.2.4.2 Coating of Catalysts Surfaces by Particulates or Coke

A second major class of processes leading to reduction in catalyst activity is the buildup of surface coatings on the catalyst. These may be either condensed (and polymerized or partially charred) organic material or layers of inorganic particulates. Their effect is to inhibit contact between the gas phase and the catalyst surface (increased mass transfer resistance). The effect on performance is similar to the curve shown for geometric area reduction in Figure 4-10 at least in the high temperature region. Raising preheat temperature is not very effective in compensating for this type of activity decline. Periodic cleaning of the catalyst is usually effective in restoring up to 90% of the initial catalyst activity where surface coating is responsible for activity loss.

A number of cleaning procedures are recommended by catalyst manufacturers for removal of surface coatings. Their recommendations should be followed, because some procedures are applicable only to specific catalysts. The following brief descriptions of cleaning procedures are included to indicate the type of maintenance required. The frequency at which such cleaning will be necessary varies greatly between applications and even between installations in a particular application. However, cleaning at three to six-month intervals appears to be a reasonable average. For organic (combustible) coatings, catalyst activity can be regenerated by raising the preheat temperature gradually to 1000 - 1100°F for 2 - 3 hours and allowing the material to burn off. Catalyst bed (or outlet temperature) should not be allowed to go over 1250 - 1300°F during this process. By removing the catalyst element, solvent washing can be used to remove the condensed organic. However, this must be done with some care, since the catalyst is active for oxidation of organic solvents and may initiate combustion of the solvent during washing.

Inorganic particulates may originate from pigment residues in metal decorating, residues of silica from silicone lubricants used in drying ovens, and corrosion products in process equipment or ducting. For inorganic materials, removal of the catalyst from the unit for cleaning is necessary. Cleaning methods include: 1) vacuum cleaning or back-blowing through the element with clean, compressed air; 2) washing the element with water or mild detergent solution (detergents containing phosphates should not be used); 3) washing and/or soaking the catalyst element in mild organic acid (e.g., oxalic acid solution) for dissolution of iron oxide deposits, followed by rinsing with water. Where tenacious deposits are not removed by such methods or where the catalyst has been exposed to high temperatures and surface coatings have fused or alloyed with the catalytic metal either replacement with new catalyst or reconditioning by the manufacturer will be necessary. Reconditioning is available only for the all-metal catalysts at a cost considerably lower than replacement with new catalyst. However, such reconditioning is possible only for certain types of catalyst deactivation.

4.2.4.3 Specific Poisoning of Catalyst Activity

The final class of deactivation processes is poisoning by specific contaminants in the waste stream which chemically combine with or alloy with the active catalyst metal. With most of the metallic poisons, the deactivation is assumed to be alloying with the active metal. The list of such metal poisons frequently quoted includes:

phosphorus
bismuth
arsenic
antimony
mercury
lead
zinc
tin

At sufficiently high temperatures, even copper and iron are capable of alloying with platinum. However, at normal temperatures below 1100°F oxides of iron only form a surface coating on the catalyst element. A recent review[4-13)] lists the first five as fast-acting poisons. Such materials, including phosphate residues from metal cleaning detergents, should be rigorously excluded from process equipment generating the waste stream. Even trace quantities of these fast-acting poisons in the waste stream can lead to rapid catalyst deactivation. The last three metals (i.e., lead, zinc and tin) are referred to in the same source as slow-acting poisons. Experience indicates that catalysts are somewhat more tolerant of these materials, particularly at temperatures below 1000°F. They probably affect the catalyst by alloying and the rate of alloying increases with temperature. These materials should also be excluded, and galvanized ductwork should not be used in the process or afterburner system.

Sulfur and halogens are also regarded as catalyst poisons. However, in most cases, their chemical interaction with the catalyst is reversible (i.e., catalyst activity is restored when the halogen or sulfur-containing species is removed from the waste stream). The high ignition and preheat temperatures listed for thiophene and chlorobutane in Table 4-2 are an indication of the inhibiting effect of sulfur and halogens on oxidation activity. Nevertheless, in at least one case, a catalytic afterburner is used in odor abatement of a stream containing butyl mercaptans as the primary odorous pollutant. Where concentrations of sulfur and halogen containing species are significant (leading to >200 ppm SO_2 or >50 ppm HCl in the afterburner effluent) scrubbing facilities downstream of the afterburner would be required to limit emissions of these species.

The effects of specific poisons on catalyst performance is qualitatively similar to that described and illustrated in Figure 4-10 for aging. With normally encountered trace concentrations of poisons in the waste stream, deactivation will be primarily uniformly distributed deactivation of chemical sites. Compensation by increasing preheat temperature will allow performance to be sustained at a satisfactory level until a large fraction of the chemical activity has been lost. Incidents of short-term exposure of the catalyst to higher concentrations of poisons may result in nearly total deactivation in a local region or throughout the bed. In such cases, catalyst replacement is the only effective way of restoring performance.

4.2.5 Users Reports on Catalytic Afterburner Performance

The response to the questionaires covering catalytic afterburners was small, only 24 units being reported. A complete tabulation is given in Chapter 11 (Table 11-2). Pertinent data have been extracted and summarized in Table 4-3. Only 30% of the units were reported to be satisfactory.

The major maintenance problem was catalyst poisoning, reported on 17 units. Other maintenance problems were negligible in comparison.

Specific data on the reduction of hydrocarbons was given by users in only one case; 90% reduction was claimed, but the unit still did not meet local requirements.

Table 4-3. USERS' INFORMATION ON CATALYTIC AFTERBURNER PERFORMANCE

Overall Performance

Number Installed	Number with Heat Exchangers	Number Not Yet Run	Performance		
			No Comment	Satisfactory	
				Number	% of Those In Use
24	7	1		7	30

4.2.6 NO_x Formation in Catalytic Afterburners

The same factors, discussed for thermal afterburners in sections 3.1.2.7 and 3.1.2.8, will contribute to detectable, but normally low, emissions of nitrogen oxides, NO_x, in the stack effluent from a catalytic afterburner. The emission factor for NO_x is expected to be the same as for a thermal afterburner, 0.05-0.1 lb NO_x emitted/MM Btu preheat fuel value required. Because of the lower level of preheat in catalytic afterburners, the concentration of NO_x in the effluent will be lower, typically about 15 ppm. As with thermal afterburners, chemically combined nitrogen in the fume stream is expected to be nearly quantitatively converted to NO_x in the effluent.

4.3 Equipment Description

4.3.1 Catalysts

The various matrix supports for afterburner catalysts have been selected or developed to provide a compact catalyst element with: 1) high geometric surface area; 2) moderately low pressure drop; 3) structural integrity and durability; and 4) a system in which uniform distribution of the flow of waste stream through the catalyst element can be accomplished readily. More conventional catalyst pellets have been used as supports for similar platinum catalysts for other applications (NO_x abatement in nitric acid plant tail gas) and are apparently used in some afterburner applications in Europe and Canada. Typically, spherical pellets or short cylindrical extrudates (L/D = 1) with a diameter of about 1/8 inch, are supported within a basket as the catalyst element in these units. To attain reasonably low pressure drops through these elements, the flow rate through the bed must be low and the element thin (1 - 2 inches) in the direction of flow. Settling of pellets within the basket and attrition of pellets lead to problems with bypassing of part of the waste stream and poor flow distribution. The matrix elements in use in this country overcome these difficulties. Commercial afterburner catalysts may be divided into two categories; the all-metal catalysts and the Pt/Al_2O_3 catalysts supported on ceramic honeycombs or other matrix elements.

4.3.1.1 All-Metal Catalysts

Only one manufacturer, UOP Air Correction Division, supplies the all-metal catalysts. These have been available and used in a wide variety of afterburner applications since 1949. In 1957, over 1000 installations using these catalysts in afterburners were reported$^{4-11)}$. Although the number of installations at present appears considerably smaller than this (based on results of the user survey made in connection with this study), a great deal of experience in use of these catalysts has been accumulated in the last twenty years. The catalyst support is made from nichrome ribbon (1/16-inch x 0.005-inch) which is crimped and formed into a mat having a void fraction of 90 - 93%v, for the manufacturer's standard "D" series catalysts. A second series of catalysts termed "Expanded", having a void fraction of 0.95 - 0.97 is offered by the manufacturer for applications requiring very low pressure drop. The active Pt or Pt/Pd component of the catalyst is applied by an electrodeposition technique to form a spongy, porous layer about 0.001 inch in thickness on the nichrome surface. The mats are mounted in stainless steel frames with heavy-gauge 16 mesh screens supporting front and back faces of the modular element. Figure 4-11 shows the manufacturer's standard "D2" module, 18" x 24" x 2-1/2", with a face area of 3 ft^2 and a volume of 0.625 ft^3. These modules weight about 25 pounds and are easily handled. Other shapes and dimensions of catalyst modules are manufactured. Most installations use the "D-2" or the similar "D-3" or "D-4" which have the same face area but are 3-3/4" and 5" thick, respectively. The catalyst modules are arranged in banks as shown in Figure 4-12. The catalyst element mounting framework is line-welded into the afterburner system, and the modular units are champed to this framework against asbestos tadpole gasketing to prevent by-passing of unreacted waste stream.

Figures 4-11 and 4-12 follow

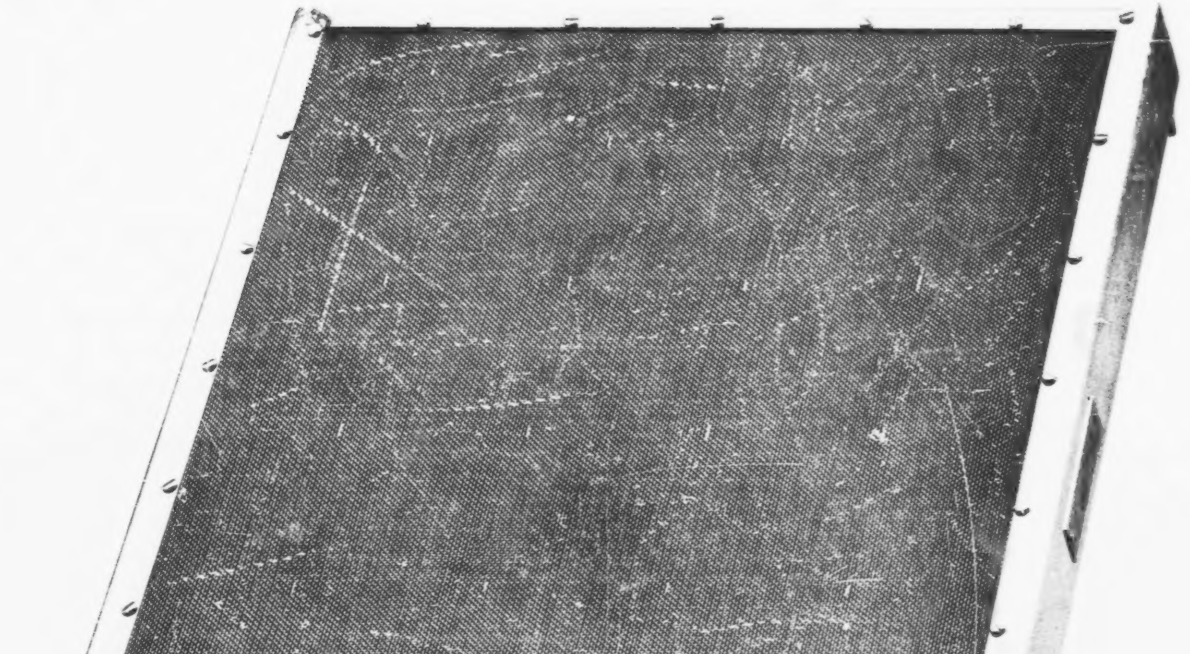

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Figure 4-11. STANDARD MODULAR ELEMENT FOR ALL-METAL CATALYST

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COURTESY: *UOP Air Correction Div.*

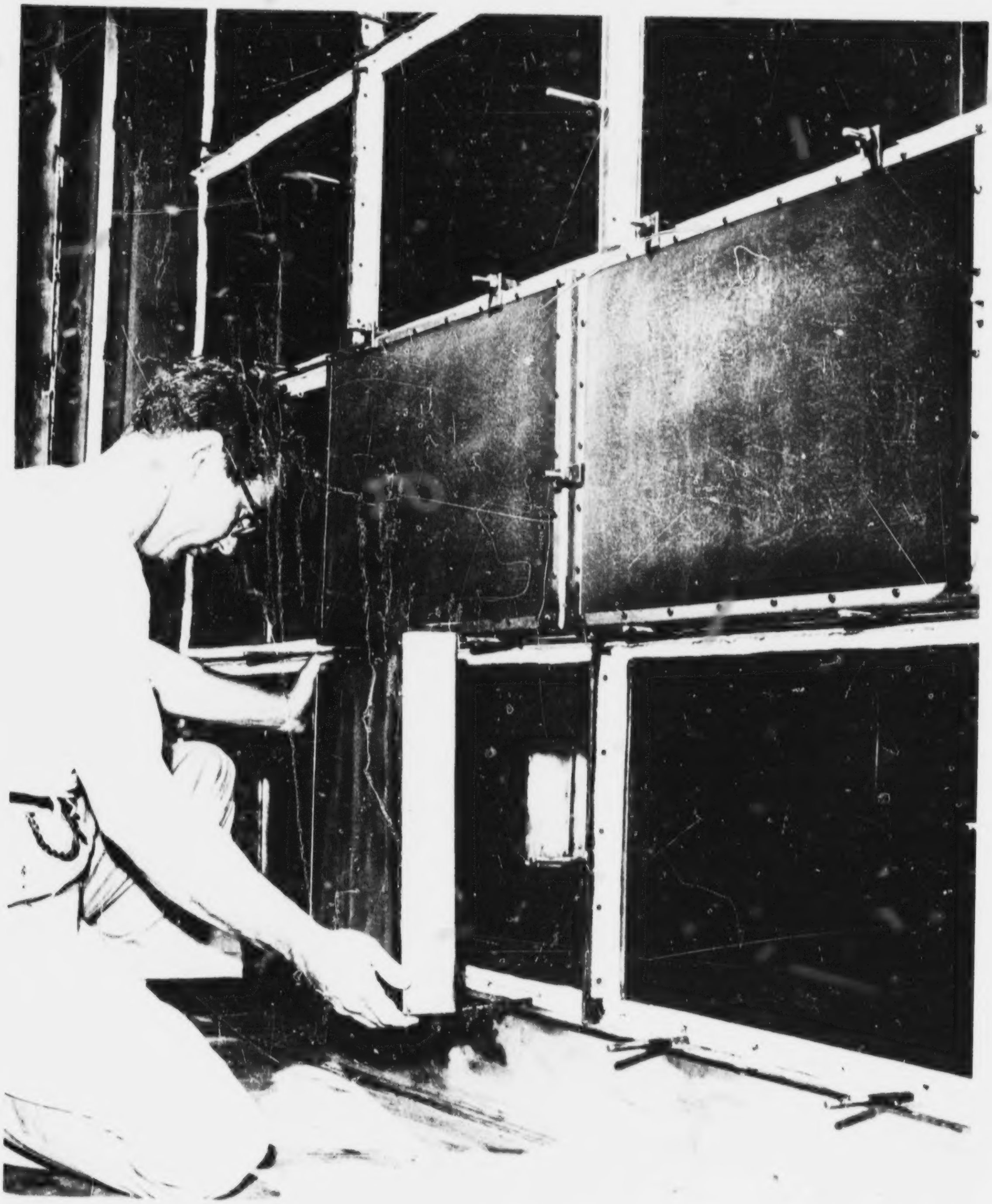

Figure 4-12. MODULAR CATALYST ELEMENTS MOUNTED IN FRAMEWORK IN AFTERBURNER SYSTEM

COURTESY: UOP Air Correction Div.

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Geometric characteristics of the UOP metal catalyst and typical design conditions for its use in an afterburner are given in Table 4-4, along with similar characteristics for the other commercial afterburner catalysts. Figure 4-13 shows pressure drop characteristics (at a typical operating temperature of 800°F) for the same group of materials. In both table and figure, 1/8-inch spherical catalyst pellets have been included for comparison. Design methods for calculation of required catalyst bed dimensions and pressure drop are discussed in Section 17.

The all-metal catalysts are rugged; the nichrome support resists oxidation and chemical attack from materials likely to be present in waste streams at temperatures up to 1500°F (the maximum exposure temperature recommended by the manufacturer). The manufacturer has developed methods for reconditioning and reactivation of these catalysts, when the deactivation or poisoning process has not damaged the support. In such cases, catalyst replacement with reconditioned elements can be considerably less expensive than replacement with new elements (see Chapter 11).

The metal base catalysts are inherently less active than the alumina base catalysts (except for a few more easily oxidized hydrocarbons) because of microstructural differences in the active metal crystallites. Consequently, designs with these catalysts usually require greater catalyst volume and/or higher preheat temperatures than do those using ceramic honeycomb substrates. However, costs per unit volume are lower for the metal catalyst, and comparisons should be made in terms of overall economics. The typical design conditions indicate low superficial velocity (face velocity) and exceptionally low pressure drop through the catalyst bed (typically less than 0.5" in H_2O through the entire bed), relative to the other catalyst configurations. The lower superficial velocity requires a larger cross section of the afterburner unit for a given throughput, as well as lower velocity in the upstream preheat/mixing section. Under similar flow conditions, the crimped-ribbon array would be expected to have a much higher efficiency for collection of particulates from the waste stream (leading to suppression of activity by surface coating) than would the honeycomb matrices. The lower operating velocity used with the metal ribbon mats partially compensates for the fact that their geometry favors particle collection.

4.3.1.2 Alumina-Base Catalysts

Alumina-base catalysts are available in three geometric types from four different manufacturers. A fourth geometric structure has also been used experimentally (including commercial scale tests). All are similar in having the basic geometric shape formed from a dense, non-porous ceramic structure (α-alumina, or one of the silica-aluminas, e.g., mullite or cordierite). On the surface of this geometric matrix, a thin, wash coat (approximately 0.005 inch in thickness) of porous α-alumina is applied. The active noble metal is deposited or dispersed within this thin layer of porous alumina as microcrystallites. Methods of application of the wash coat and dispersion of active metal within it vary among manufacturers and techniques for these processes are either patented or considered proprietary. Activities of the finished catalysts vary somewhat between manufacturers but all exhibit similar chemical oxidation characteristics. For

Table 4- 4. PROPERTIES AND TYPICAL DESIGN PARAMETERS FOR AFTERBURNER CATALYSTS

Catalyst Type	Metal Ribbon	Oxycat	1/8 inch Hex-Cell Honeycomb	8C/inch Corrugated Honeycomb
Manufacturer(s)	UOP	Oxycatalyst	Oxycatalyst "Oxycomb" Matthey-Bishop "THT" duPont	Engelhard
Geometric Surface Area (ft_2/ft_3)	330	36	270	700
Density (lb/ft^3)	40	50	40	40
Cost ($\$/ft^3$)	400 - 500	480	1300 - 1800	2200 - 2500
Design Conditions for 90% Conversion of Typical Solvent Hydrocarbon at 10% LEL				
Catalyst volume (ft^3/M SCFM waste)	2	5 - 9	1 - 1.2	0.6 - 1
Bed length (inch)	2.5 - 5	9 - 18	8 - 10	2 - 6
Superficial gas velocity				
($SCFM/ft^2$)	150 - 200	200 - 400	600 - 1000	200 - 400
(ft/sec at preheat temp.)	6 - 8	8 - 15	25 - 35	8 - 15
Temperature (°F)				
Preheat	800	650 - 700	650 - 700	650 - 700
Outlet	1050	900 - 950	900 - 950	900 - 950
ΔP (in H_2O)	0.3 - 0.5	0.5 - 1.5	2 - 4	1.5 - 3

Figures 4-13 and 4-14 follow

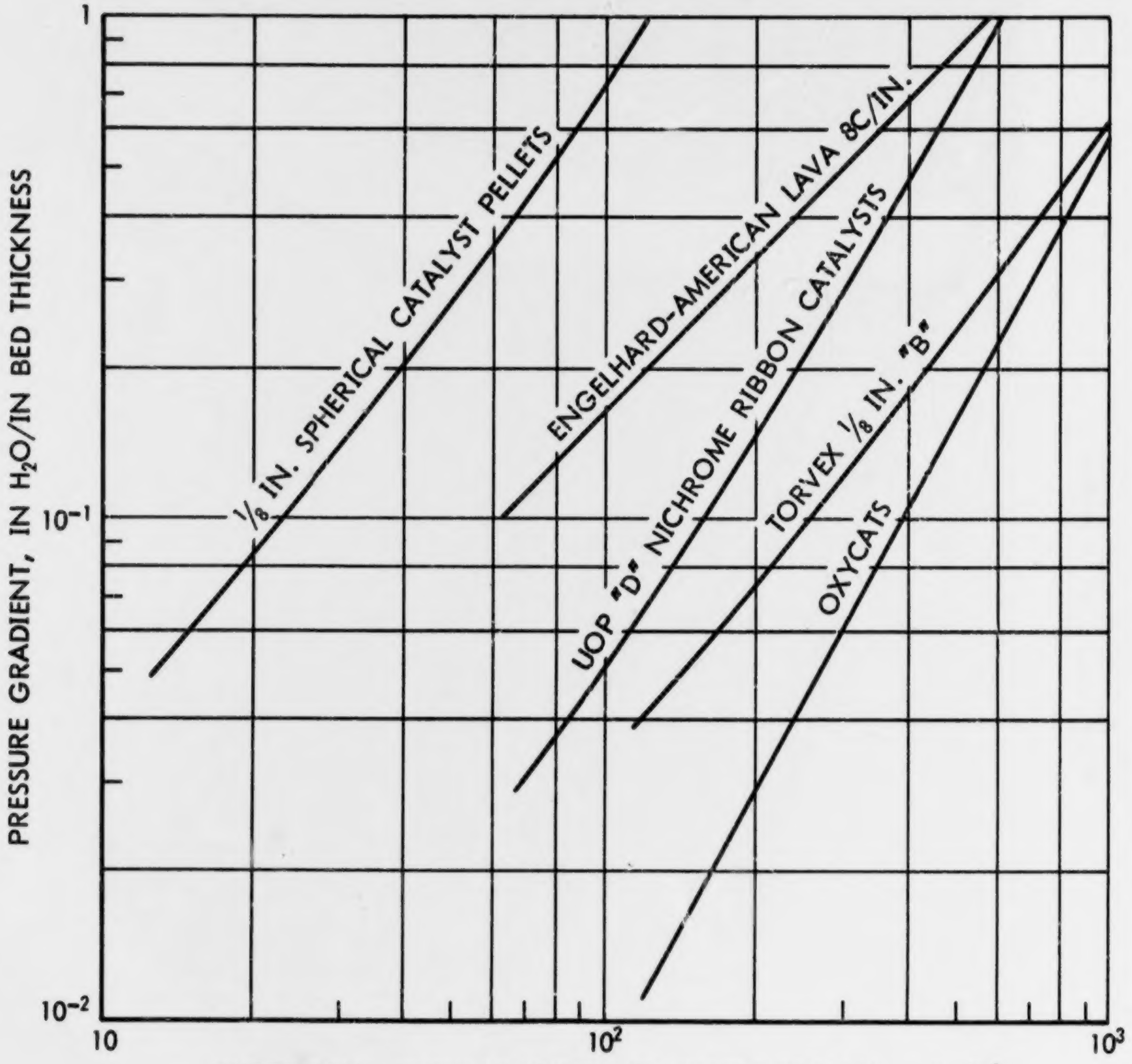

Figure 4-13. PRESSURE GRADIENTS FOR FLOW THROUGH CATALYST MATRICES AND BEDS AT 800°F, ATMOSPHERIC PRESSURE

76h

Oxycats   Torvex 1/8 in "B"
UOP "D" Nichrome Ribbon Catalysts
Engelhard-American Lava 8C/in
1/8 in Spherical Catalyst Pellets

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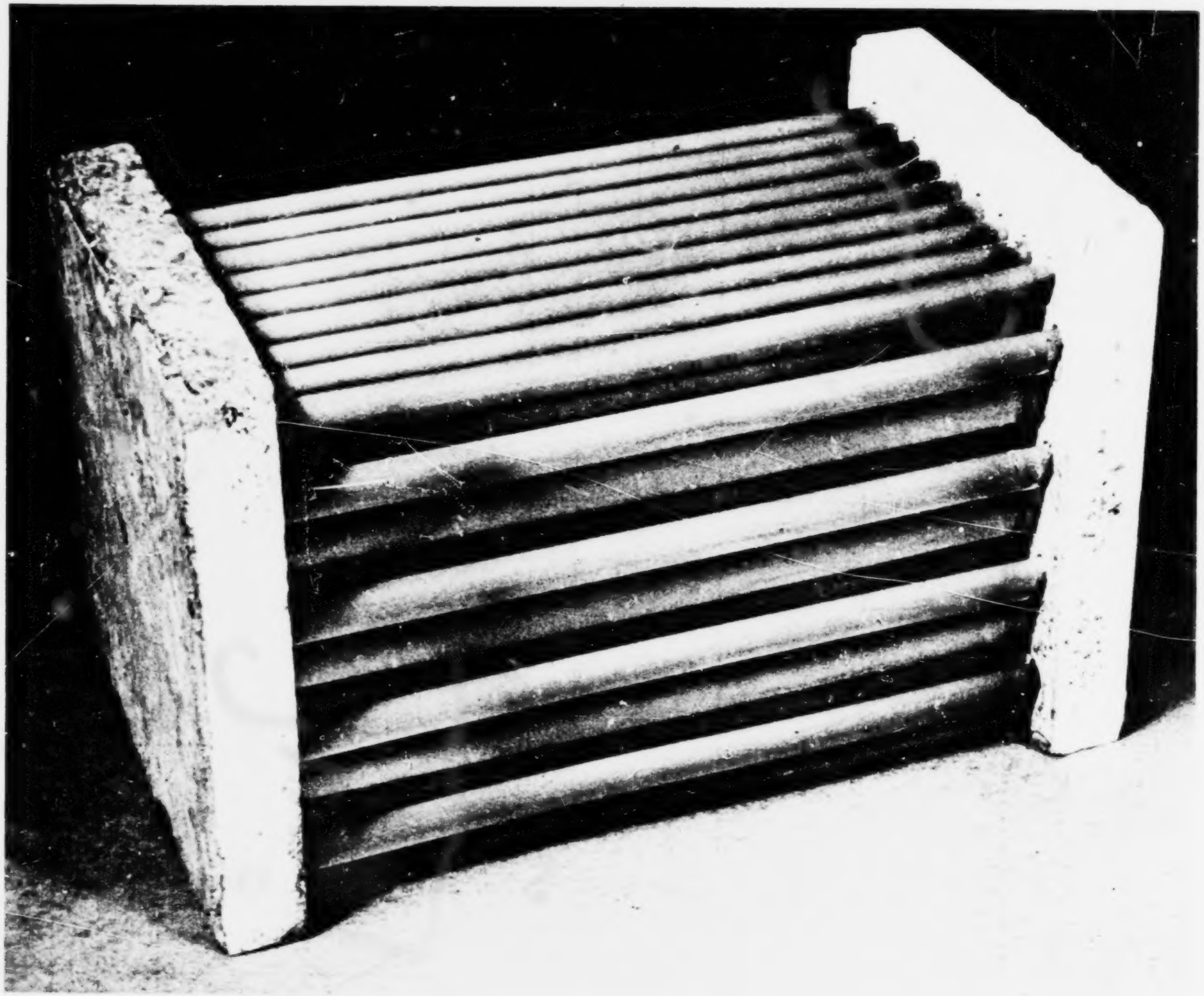

Figure 4-14. OXYCAT

5½" Long x 3 3/16" Wide x 3" Deep

76c

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COURTESY: *Oxy-Catalyst, Inc.*

many materials they are significantly more active in promoting oxidation than are the all-metal catalysts. Each manufacturer may have several formulations for noble metal content and type, and methods for preparing catalysts with special activity for particular waste stream components.

The ceramic matrix structure longest in use is the "Oxycat" (Oxycatalyst, Inc.). The basic element is a grouping of ceramic rods, arranged with end plates into a small, brick-like structure 5-1/2 inches long x 3-3/16 inches wide x 3 inches deep (in the direction of flow) shown in Figure 4-14. The individual rods are of neutral air foil cross section with blunt ends facing the oncoming gas flow. These have an effective diametral dimension of about 0.15 - 0.2 inch and have the porous alumina/noble metal dispersion on their external surface. These modular bricks are arranged into a wall or bed usually 3 to 6 bricks deep in the direction of flow (which may be either vertical or horizontal). Channel connectors may be used to hold adjacent rows of "Oxycats" together in this type of bed. The flow of preheated waste stream through this array is then similar to cross flow through a staggered tube-bundle.

Properties and normal design range for Oxycats are given in Table 4-4 and estimated pressure drop characteristics in Figure 4-13. Design methods are discussed in Chapter 17. Costs are given in Chapter 11.

As shown in Table 4-4, the surface area per unit volume of this structure is much lower than for the other matrix supports. Consequently the catalyst bed would need to be 5 - 15 times the volume of Oxycats as of the other types of catalyst. This results in an afterburner unit much larger and heavier when Oxycats are used. The manufacturer, Oxycatalyst, Inc., now uses "Oxycomb" catalysts (based on 1/8-inch hex-cell honeycomb ceramic) for most new installations. Oxycats are still manufactured for replacement in existing units and have been used in some new installations where very low pressure drop was required.

Catalysts prepared on two types of ceramic honeycombs are commercially available and in use.

A 1/8-inch hex-cell honeycomb, "TORVEX-B" (manufactured by E. I. duPont Company) is used as a base for "Oxycomb" (Oxycatalyst, Inc.), "THT" (Matthey-Bishop) and catalysts manufactured by duPont. Figure 4-15 shows the structure of this honeycomb ceramic (in several hole sizes). The honeycomb support is available in 12" x 12" x 1" thick sheets. Catalysts prepared on these sheets are used to assemble modular catalyst elements or total catalyst beds for incorporation into afterburners. Typical modules may be 4 - 13 inches (sheets) thick. Manufacturers differ in their design of catalyst module or catalyst bed enclosures, but all are designed to facilitate removal of the catalyst for cleaning or replacement. Figure 4-16 shows an exploded arrangement of catalyst, catalyst module frame, and a section of mounting framework. The module shown is 12" x 12" x 3". The photo shows the module made up with an experimental "cross-flow" honeycomb rather than the hex-cell type. The overlayed sheets of honeycomb catalyst are first framed with a coating of compressible, castable ceramic. This inner element then fits snugly into an outer stainless steel frame which can be mounted in the framework in the afterburner system.

Other manufacturers include retaining screens and structural supporting members at front and back sides of the metal frame of the catalyst element. Figure 4-17 shows such an element where the entire catalyst charge for an afterburner unit has been made up as a single element. The 12" x 12" modules may weight 20 - 50 pounds each, while the total bed element shown might have a total weight of 500 pounds for a moderate size unit.

The significant design and construction features for honeycomb ceramic catalyst modules and mountings are related to providing uniform flow of waste stream through the catalyst and avoiding by-passing of the bed by any part of this stream. As with the all-metal catalysts, the mounting framework should be line welded or otherwise tightly sealed to the afterburner enclosure walls. Catalyst elements should be tightly seated and sealed in mounting frameworks. With normal pressure drops of 3 - 5 inches across the catalyst element, even small openings around the catalyst could lead to significant by-passing. Mounting frames are designed to seal around the ceramic honeycomb without placing it under strain during thermal cycling (e.g., startup or shutdown of the afterburner). Some cracks do develop in these catalysts in use. However, they are prevented from opening to significant size by the element mounting frames.

A similar but geometrically and dimensionally different honeycomb structure is used in catalysts manufactured by Engelhard Industries. The ceramic matrix, "Thermacomb" (American Lava Division of 3M Company) is of corrugated construction with passageways for gas flow more nearly triangular than circular in cross section (see Figure 4-18). In current afterburner catalysts the eight corrugation/inch size of the honeycomb material is used. Gas flow passages have a dimension of 0.05 - 0.06 inches. This material has the highest surface to volume ratio of any of the matrix materials used. Modules made up from this material are shown in Figure 4-19.

Properties and typical design conditions for ceramic honeycomb catalysts are given in Table 4-4, and pressure drop characteristics in Figure 4-13. Design methods are discussed in Chapter 17. Costs are given in Chapter 11.

An additional honeycomb structure, "TORVEX-C", has been used on an experimental basis as a support for afterburner catalysts. This material, shown in Figure 4-20 has flow passages directing gas flow at 45° to the main flow direction. It has a significantly higher pressure drop than "TORVEX-B" honeycombs, and its performance as a catalyst support does not differ greatly from "TORVEX-B". Data available on performance are inadequate to evaluate it for this use at present. It has been used in connection with the "TORVEX-B" honeycombs as an inlet distribution layer, and a commercial trial has been made in a paint baking application.

Figures 4-15 through 4-20 follow

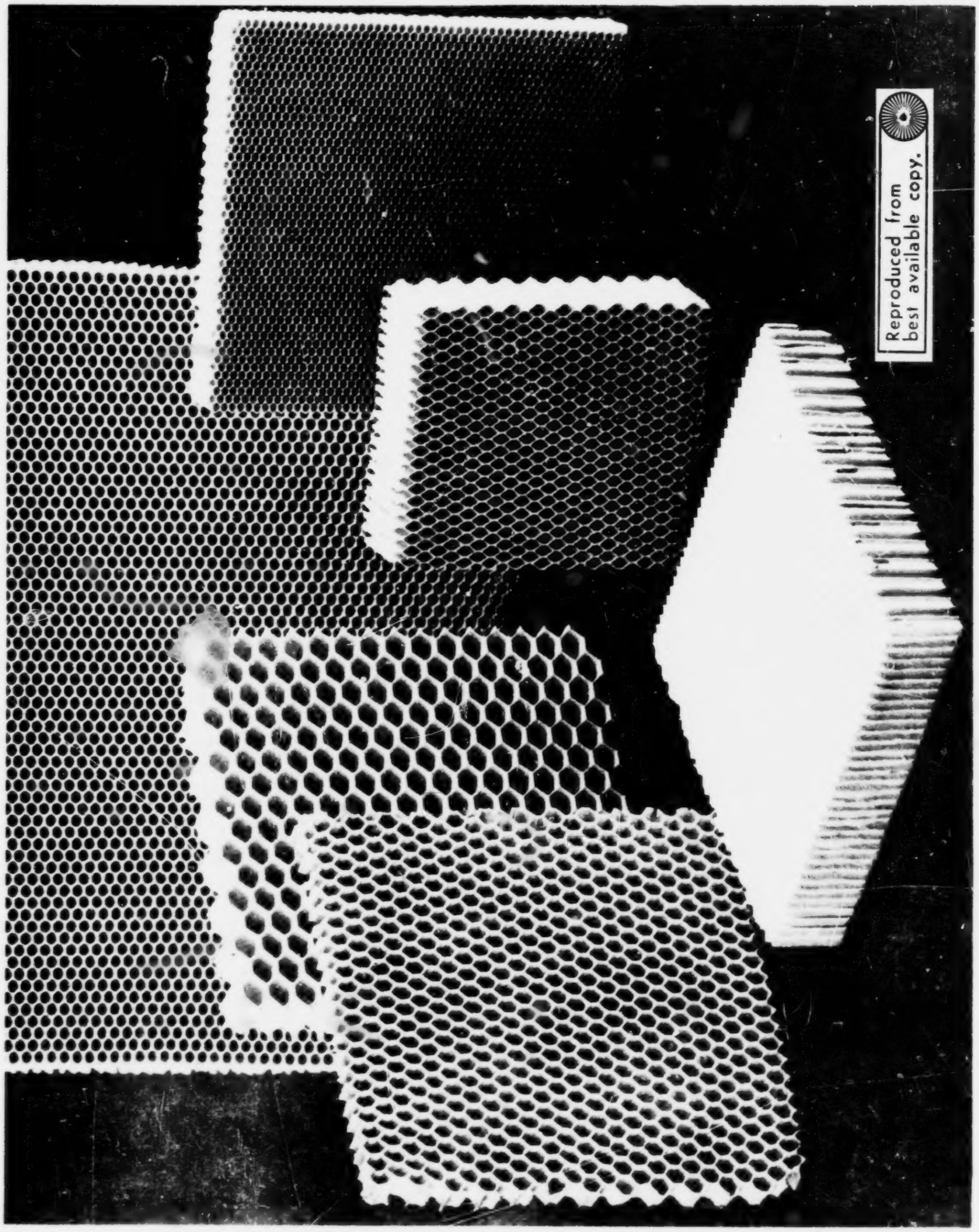

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Figure 4-15. "TORVEX B" HONEYCOMB ALUMINA CATALYST SUPPORT

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COURTESY *duPont, Industrial and Biochemicals Dept.*

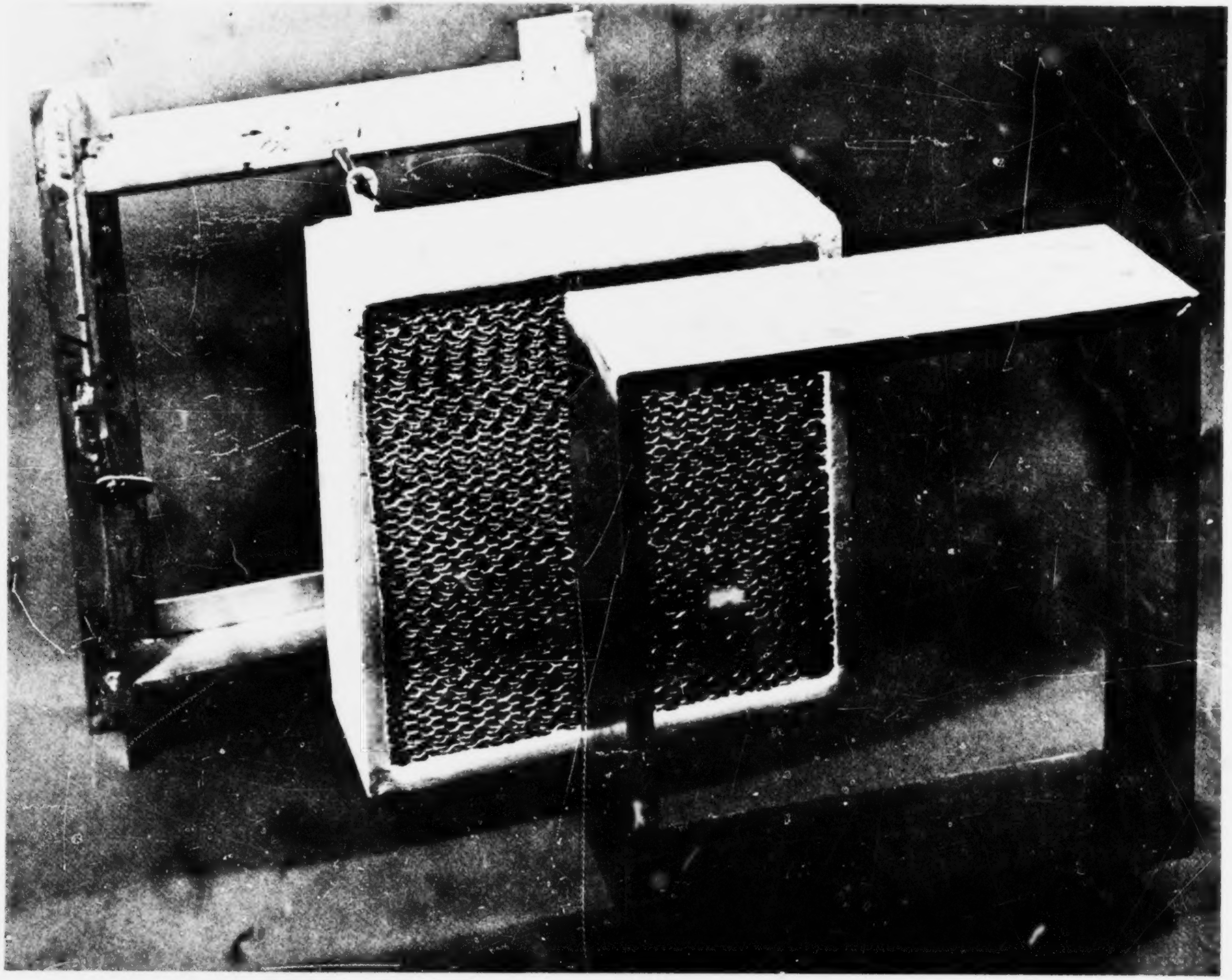

Figure 4-16. EXPLODED VIEW OF HONEYCOMB CATALYST MODULE

7

COURTESY: duPont, Industrial and Biochemicals Dept.

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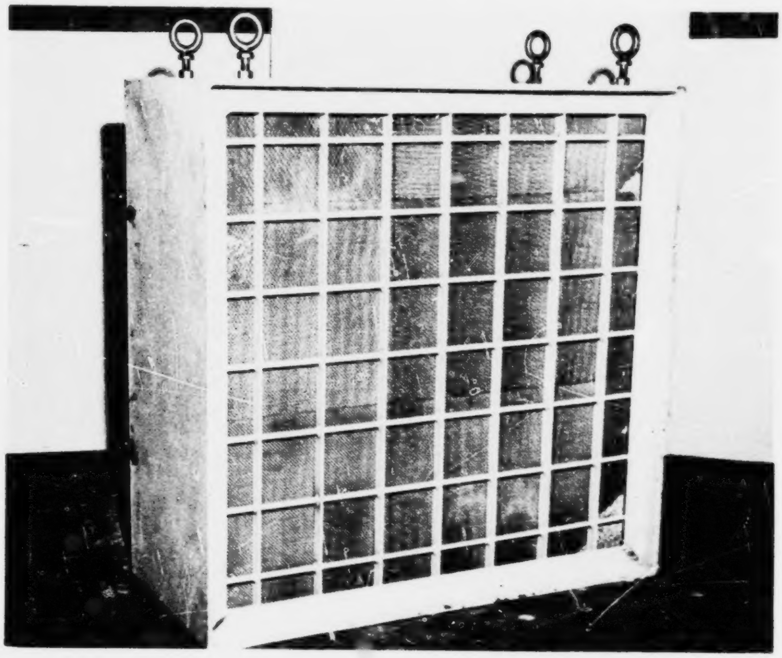

Figure 4-17. CERAMIC HONEYCOMB CATALYST ELEMENT
Single Element Contains Entire Catalyst Charge for Afterburner

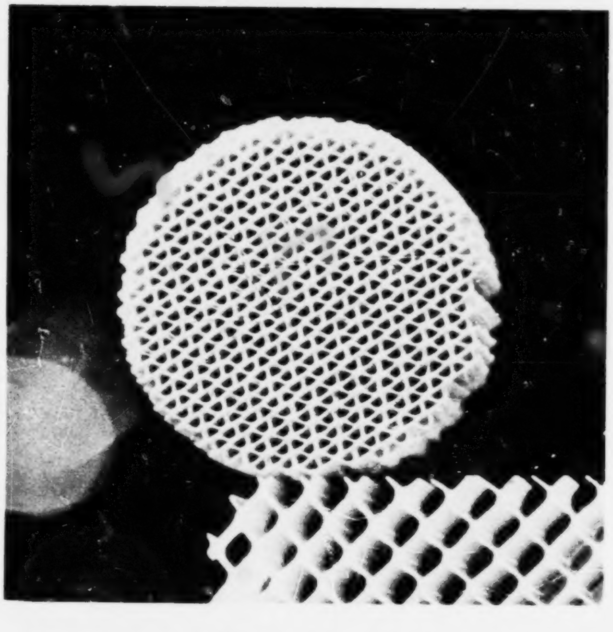

Figure 4-18. "THERMA COMB" HONEYCOMB CATALYST SUPPORT

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COURTESY: *Oxy-Catalyst, Inc.*

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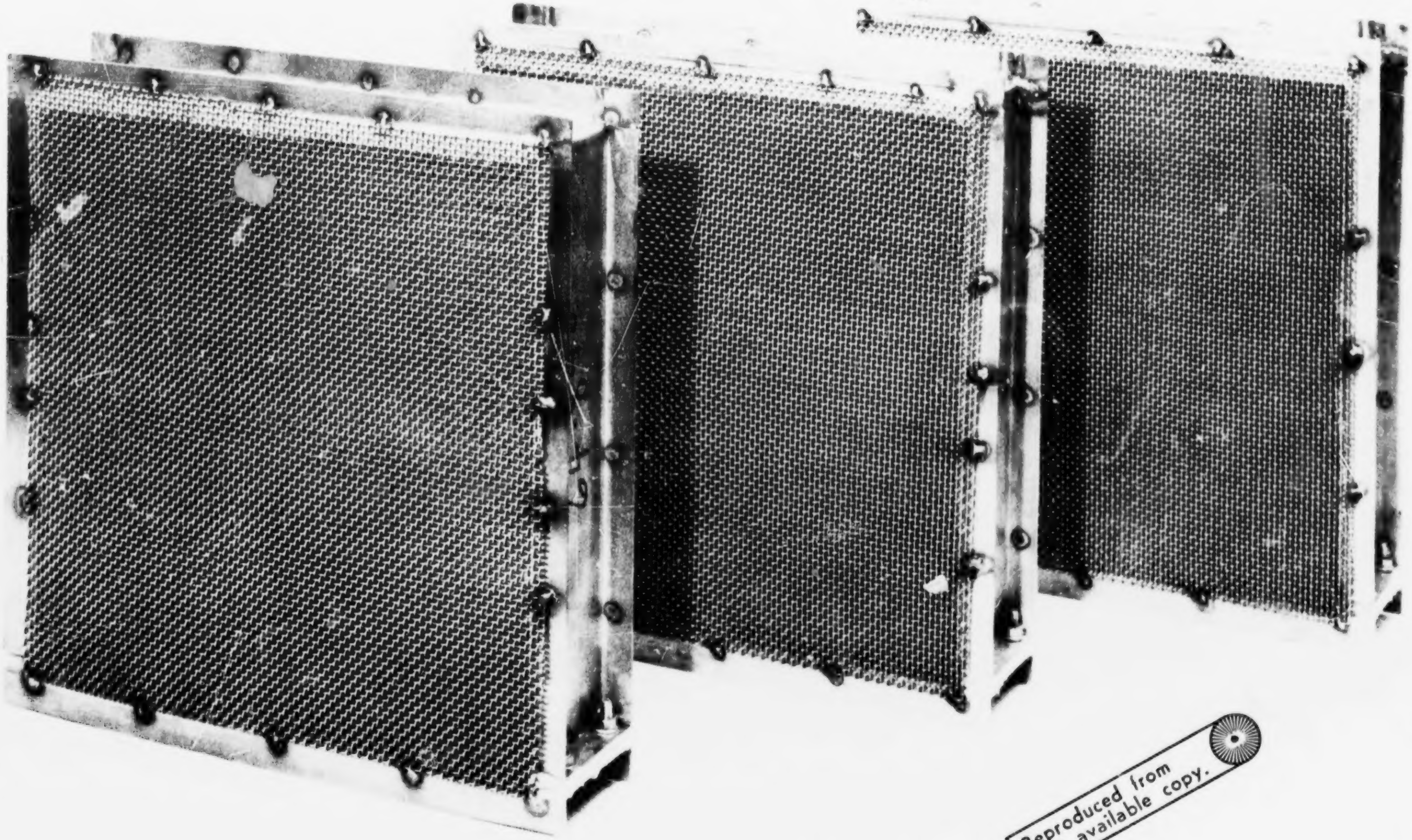

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Figure 4-19. CATALYST MODULES MADE FROM THERMA COMB 8C/INCH

COURTESY: *Engelhard Industries, Gas Equipment Section*

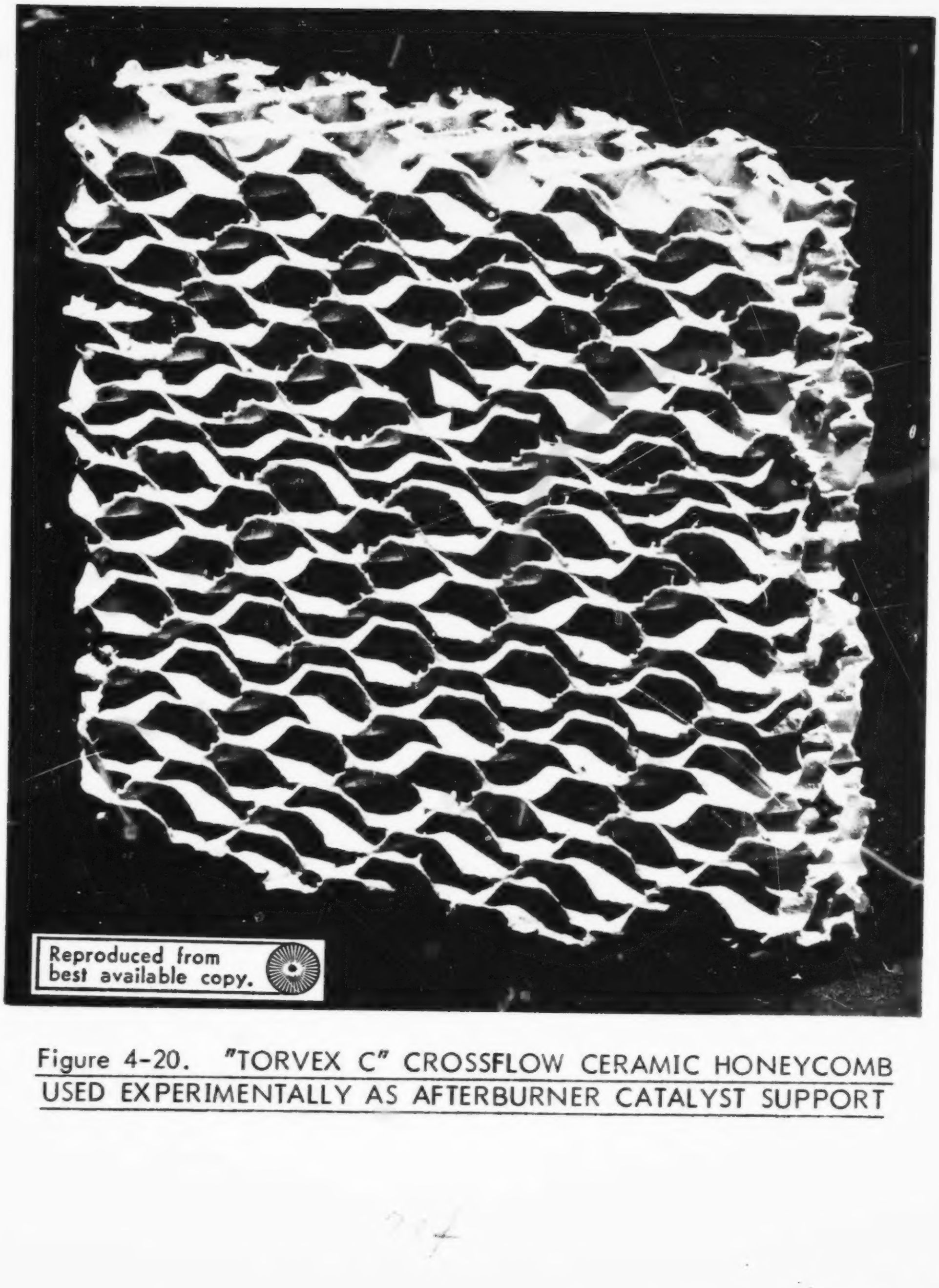

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Figure 4-20. "TORVEX C" CROSSFLOW CERAMIC HONEYCOMB USED EXPERIMENTALLY AS AFTERBURNER CATALYST SUPPORT

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COURTESY: *duPont, Industrial and Biochemicals Dept.*

4.3.2 Catalytic Afterburner Enclosures

The enclosure for a catalytic afterburner unit provides a duct in which the catalyst bed or matrix is mounted. Upstream of the bed, the enclosure provides means of preheating the waste stream to be treated, mixing the waste stream with preheat burner combustion products, and distributing uniformly the flow of preheated waste stream through the catalyst. The system also includes appropriate instrumentation for measuring and controlling temperature of the catalyst bed and for operation of the preheat burner. It may include a fan to overcome pressure drop through the afterburner if the process exhaust fan does not have sufficient head to overcome this pressure drop. Downstream of the catalyst the system may include provision for heat recovery, or exhaust directly to a (usually short) stack.

4.3.2.1 Burners

All of the known installed catalytic afterburner systems use natural gas as fuel for preheat burners. Some designs using oil-fired burners have been proposed by manufacturers for specific installations. In such units oil burners have the usual disadvantages vis-a-vis gas burners, e.g., lower turndown ratio, primary air required (normally from external source) increases the total flow through the catalyst, and the single discrete oil burner contributes far less to mixing of the fume with combustion products than does the distributed gas burner. In addition, the sulfur in the fuel (oil) may exert a significant depressing or poisoning effect on catalyst activity, requiring the use of more catalyst or higher temperatures in an oil fired unit. Even No. 2 fuel oil normally has a sulfur content of 0.1 - 0.5%w and may exceed 1.0%w. Thus, where possible, gas-fired preheat burners are used.

Nearly all recent designs and installations of catalytic afterburners utilize distributed raw-gas burners (e.g., Maxon "Combustifume", Eclipse "AH", North American "Flame Grid",Pyronics or UOP burners). With such a burner, mixing of the waste stream and preheat-burner combustion-products is accomplished rapidly downstream of the burner due to design features of the burner (see Section 3.2). The use of a raw-gas burner (using the waste stream as combustion air) also has the advantage of requiring no additional combustion air and hence a lower total flow through the catalyst bed. (With external primary air to the burner, the flow through the catalyst would be increased perhaps 15 - 25%, depending on the percent excess air used). As explained in Section 3.2, such use of a raw gas burner requires a minimum oxygen content in the waste stream of about 15 - 16%v. However, in most applications the waste stream either has an oxygen content greater than this minimum, of has been air diluted (for safety reasons) to such an oxygen content prior to reaching the afterburner.

In catalytic afterburner installations, distributed burners are usually installed with blanking or "profiling" around the burner shape (see Section 3.2.1.1) such that the pressure drop across the burner is 0.5 - 2.0 in H_2O. While this practice may contribute significantly to the total pressure drop through the afterburner, it is necessary to obtain good mixing and flow distribution downstream of the burner.

4.3.2.2 Enclosure Configurations

Figure 4-21 shows a configuration much described in the literature for a catalytic afterburner enclosure. Although this design is not currently manufactured, many units of this type are still successfully in use in various applications. This configuration was developed before the availability of distributed burners, and hence relies on design features of the enclosure to accomplish mixing between preheat burner combustion products and waste stream. The burner used is a discrete "torch" type with its own primary air supply. The waste gas stream is brought in above and upstream of the burner. Two means are provided to promote mixing. The combined stream is passed through a fan located between burner and catalyst. Fluid mechanical mixing within the duct was promoted by raising duct velocity to the maximum possible and by arranging flow around a U-bend in the chamber. However, as noted in Chapters 3 and 14 neither of these techniques is very effective at providing good mixing. The configuration requires a fan rated for high temperature use (900°F). To reduce unit size and allow higher velocity through the duct, than was permissible through the catalyst element, the latter was mounted at an angle to the duct flow. In some applications, these mixing provisions were inadequate and stratified flow through the fan occurred, or non-uniform distribution of flow through the catalyst element resulted. Thus, the enclosure or duct design features contributing to gas mixing and flow distribution in this design cannot be said to have been successful in all applications. In some more recent designs, perforated metal plates have been placed between the discrete preheat burner and the catalyst, to promote mixing. A significant pressure drop (~1 in H_2O) must be taken across such a plate in order for it to contribute effectively to mixing.

With the development and availability of distributed raw-gas burners, simpler configurations for catalytic afterburner enclosures have been used. In these, mixing of waste stream and preheat-burner combustion products is accomplished rapidly downstream of the burner due to design features of the burner (see Section 3.2). The enclosure then becomes a straight duct with a uniform cross section leading from burner to catalyst, as shown in Figure 4-22. Figure 4-23 shows an external view of a completed installation of this type of catalytic unit.

The duct cross section may be either circular or rectangular (as shown in the figure) and is sized to provide an appropriate velocity of the preheated stream through the catalyst bed which is mounted normal to the flow and fills the duct. The cross-sectional area of the enclosure will be directly proportional to the throughput (including preheat combustion products) and inversely proportional to the flow velocity. Typical design velocities for gas flow through the enclosure duct are given in Table 4-4. These vary considerably with the type of catalyst matrix being used in the unit. Thus, cross-sectional areas for enclosures for normal designs with 1/8-inch Hex-cell honeycomb catalysts will normally be smaller (by a factor of 2 - 5) than for designs with the other catalyst matrix types.

The length of the enclosure between preheat burner and catalyst is normally in the range of 5 - 10 feet, independent of the enclosure cross section (which is determined by unit throughput). This length is selected to avoid overheating of the upstream face of the catalyst element due to flame impingement or

Figures 4-21 through 4-23 follow

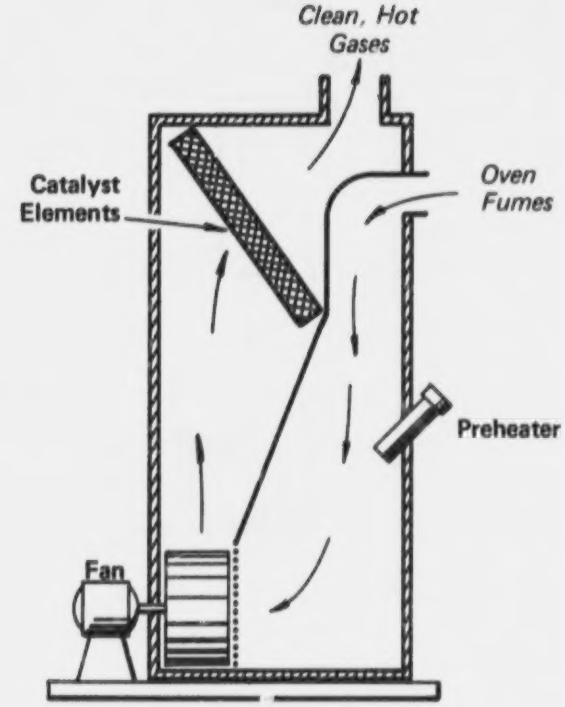
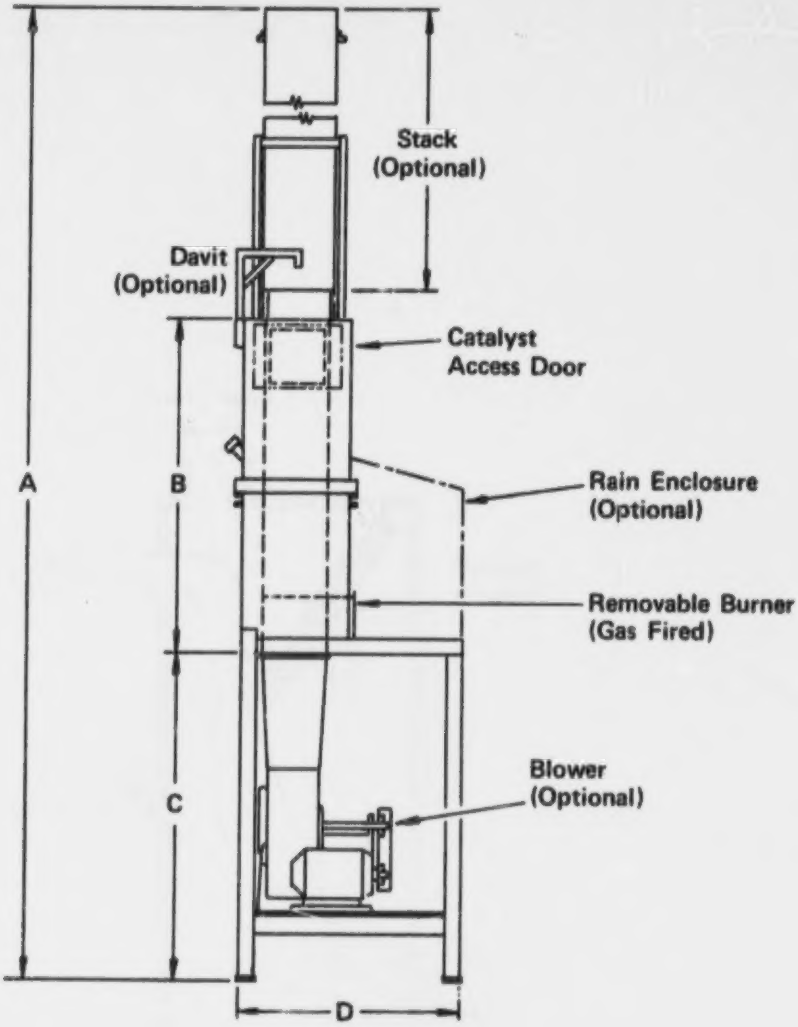
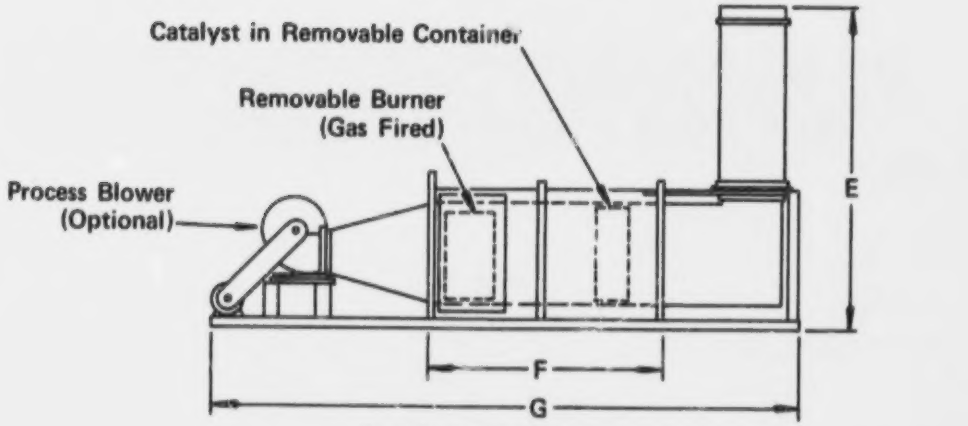

Clean, Hot Gases
Catalyst Elements
Oven Fumes
Preheater
Fan

Figure 4-21. SCHEMATIC DIAGRAM OF CATALYTIC AFTERBURNER USING TORCH-TYPE PREHEAT BURNER WITH FLOW OF PREHEATED WASTE STREAM THROUGH FAN TO PROMOTE MIXING


Stack
(Optional)
Davit
(Optional)
Catalyst
Access Door
Rain Enclosure
(Optional)
Removable Burner
(Gas Fired)
Blower
(Optional)
A
B
C
D

A. Vertical Mounting


Catalyst in Removable Container
Removable Burner
(Gas Fired)
Process Blower
(Optional)
E
F
G

B. Horizontal Mounting

Figure 4-22. IN-LINE ARRANGEMENT OF PREHEAT BURNER AND CATALYST ELEMENT

80a

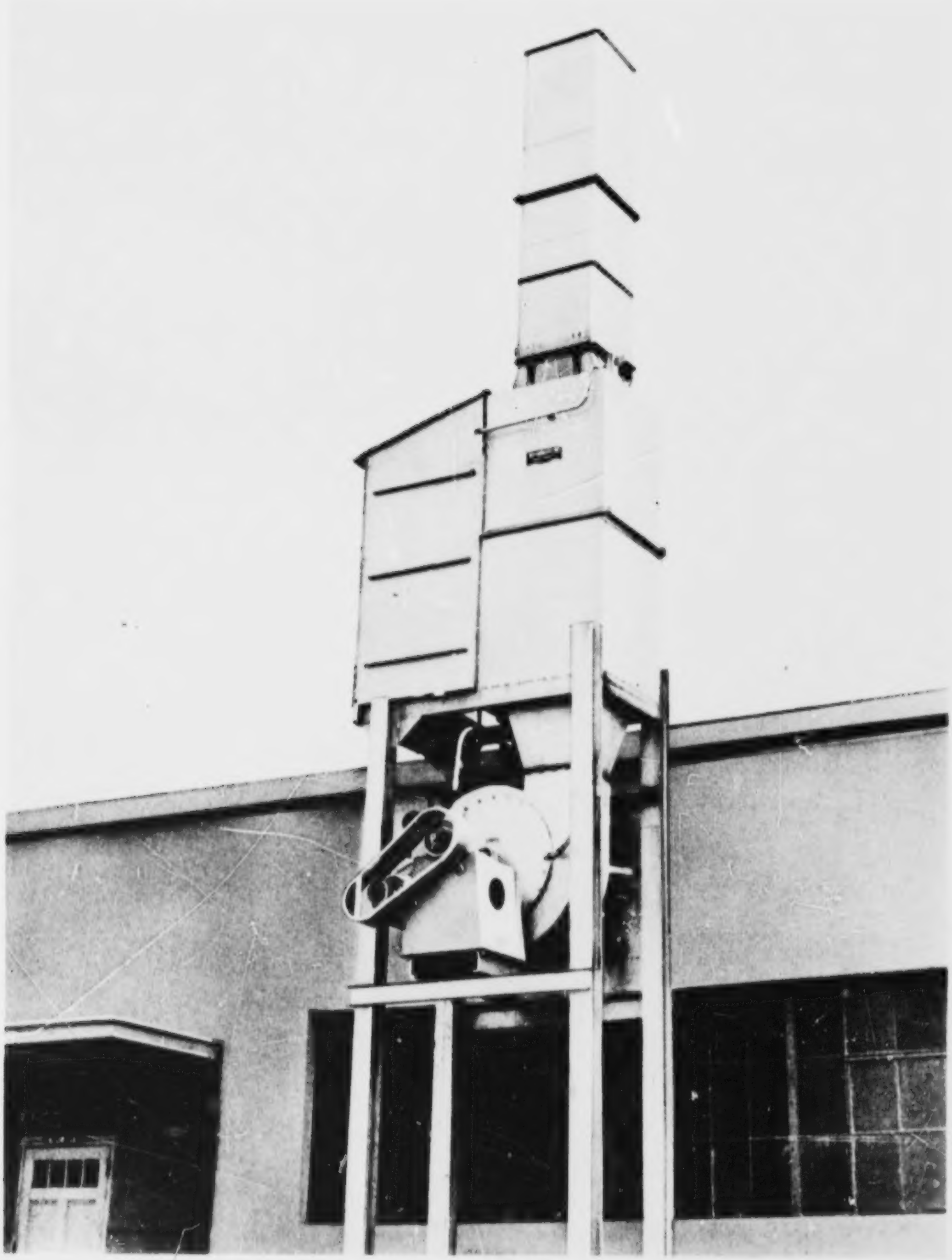


Figure 4-23. EXTERIOR INSTALLATION OF CATALYTIC AFTERBURNER UNIT

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COURTESY: *Oxy-Catalyst, Inc.*

excessive radiation from the preheat burner. The distance required to avoid these factors varies with the particular burner being used. New designs and installations have been satisfactory from this standpoint. However, in at least one installation where an older unit has been modernized by installation of a distributed burner, overheating of the catalyst by location of the burner too close to it is believed responsible for relatively short catalyst life experienced after the modernization.

In horizontally mounted units, some extension of the enclosure downstream of the catalyst element, before turning into a vertical stack is desirable, a minimum distance of 2 - 3 effective duct diameters is usually necessary to avoid maldistribution of gas flow through the catalyst due to effects of the flow around the bend into the vertical discharge. In vertically mounted units, a short discharge stack is usually provided to direct hot gas discharge and protect the downstream face of the catalyst from cooling by excessive radiation.

Extension of the enclosure upstream of the preheat burner is essentially inlet ducting and plenum which is subjected to relatively mild temperature conditions (400 - 600°F even when a recuperative heat exchanger is used). Except in those older designs where the fan was placed between preheat burner and catalyst, the fan is placed upstream of the afterburner unit and the afterburner operates under (slight) positive pressure. The overall pressure drop through catalytic units (including preheat burner and catalyst) is normally in the range of 2 - 5 in H_2O; or when a tubular, recuperative heat exchanger is included in the system 6 - 10 in H_2O.

At least two equipment manufacturers are currently offering afterburner systems which can be used either as catalytic or thermal systems. Presumably the system would be installed and operated initially with catalyst in place. If operation in this manner were satisfactory and adequate catalyst activity obtained, catalytic operation would be continued. If such operation did not prove successful, or if local performance regulations were made more stringent, such that they could not be met by catalytic operation, the catalyst could be removed and the temperature of operation increased to operate the unit thermally. Such a dual-purpose design requires that the enclosure meet all the normal requirements (e.g., temperature capability and residence time) for thermal operation in addition to providing for catalyst mounting and even distribution of flow through the catalyst matrix. Such units would have purchase costs somewhat greater than, and installation costs about equal to, thermal units. However, once installed they would provide the guarantee of performance of a thermal unit, and the potentiality of the fuel savings attainable with a catalytic unit. With use of 1/8-inch Hex-cell ceramic base catalyst, the normal design superficial velocity through the catalyst (25 - 35 ft/sec) is in the range satisfactory for thermal operation. Thus design of the enclosure to meet normal thermal design requirements would only involve making the length adequate for about 0.5 second residence time at 1400 - 1500°F and the use of materials capable of withstanding the higher temperatures of thermal operation. With the other catalyst supports, the design of a dual purpose enclosure requires mounting of catalyst other than normal to the flow to obtain sufficiently low velocity through the catalyst for normal catalytic operation.

4.3.2.3 Materials of Construction

Catalytic afterburner enclosures are subject to considerably less severe conditions than are thermal units. Consequently, somewhat less expensive and lighter weight materials and techniques can be used in their construction. Two general techniques of enclosure wall construction are used by manufacturers; 1) an outer steel structural shell lined with castable refractory (similar to construction of some thermal afterburners; 2) double wall construction (similar to the walls of paint baking or food drying ovens) in which a metal liner forms the interior wall of the enclosure. This is backed by lightweight block or mineral wool insulation, covered on the outside with a second metal shell.

Units of the first type normally have a shell of 3/16-inch thick carbon steel plate, with 4-1/2 - 5 inches of lightweight castable refractory lining. These units weight 4200 to 8500 pounds for capacities of 2500 - 12,000 SCFM waste stream. Units in this size range are shipped pre-piped and pre-wired to minimize field installation costs. Even afterburners of this construction up to 20,000 SCFM are shipped as single, pre-assembled units.

The second (double-wall) construction method is used by a number of manufacturers, with considerable variation among them on metal types and thicknesses used. The metal inner liner is generally 1/16 inch or No. 16 gauge metal either of stainless steel or aluminized steel. The range of temperatures encountered in catalytic afterburners makes the use of aluminized steel marginal for this application (see Section 7.3.1.3). In the preheat mixing section average temperatures are normally 600 - 950°F. However, local regions may run higher, and temperatures downstream of the catalyst may be as high as 1100 - 1150°F with preheat raised to obtain satisfactory performance with a partially deactivated catalyst. Furthermore, aluminized steel would have little margin of safety in the event of a control failure which allowed operating temperature to rise above the normal range. Instances of failure of aluminized steel liners have been reported in the survey of users made for this study. Insulation between walls of the enclosure is generally 3 - 4 inches thickness with type and temperature capability selected to withstand 1200 - 1500°F next to the inner liner, and to provide sufficiently low temperatures for personnel protection at the outer wall. In adequately designed units, the outer wall provides a substantial structure for the enclosure, with 1/8 - 3/16-inch carbon steel plate used. One manufacturer's designs, have, however, specified construction with No. 16 gauge inner shell and No. 20 gauge outer shell. This construction is not considered adequate for a durable installation.

Internal support frameworks for catalyst modules are usually constructed of a stainless steel or other alloy capable of withstanding temperatures of 1400°F. Construction of these parts with lesser temperature capabilities is not recommended.

4.3.2.4 Instrumentation

Primary operating control of a catalytic afterburner normally uses the outlet temperature (downstream) from the catalyst element to adjust fuel to the preheat burner. Most manufacturers provide a thermocouple to sense this temperature and an indicating (or recording) proportioning controller to actuate the fuel valve. A second thermocouple in this downstream location and thermocouple upstream of the catalyst bed (which senses the temperature of the preheated stream) are connected to a temperature indicator (or recorder) and high temperature limit instrument. This functions to prevent overheating the catalyst or afterburner in the event of fuel controller failure or sudden surges of high concentration of hydrocarbons in the waste stream. The temperature difference across the catalyst bed gives an indication of the activity of the catalyst and provides a basis for (manually) resetting the outlet (control) temperature or for making decisions regarding cleaning or replacement of catalyst (see Section 4.1.3.3). The desirability of development of a more direct means of measuring catalyst performance was discussed in that section. However, at present no inexpensive instrument exists for that purpose.

Safety instrumentation corresponding to FIA, FM, or other insurance code specified by the customer is normally included in a catalytic afterburner installation (see Chapter 9).

Pressure taps (and in some cases draft gauges) are provided for measurement of pressure drop across the catalyst bed and across the preheat burner. Excessive pressure drop across the catalyst bed is usually caused by plugging of the matrix element by buildup of particulate matter, and indicates the need to remove and clean the catalyst.

Frequent preventative maintenance of electromechanical controls is recommended for catalytic as well as thermal systems. Although thermocouples are not subjected to as high temperatures in catalytic systems, they should nevertheless be periodically tested and/or replaced. The exposure of the catalyst to excessive temperature (deactivating it) in the event of thermocouple or control instrument failure can be a costly reminder of the need for routine maintenance on these control elements.

Chapter 5. USE OF FIRED PROCESS HEATERS AND UTILITY FURNACES AS AFTERBURNERS

5.1 When To Use As Afterburners

When waste fume streams occur in an operation which is part of a plant complex containing process heaters, utility furnaces, or steam generating boilers, it may be possible to use these units as afterburners. Most such heaters operate with firebox temperatures in excess of 1800°F and with residence times in the firebox of 0.5 to 3 seconds, so that they might well serve as efficient fume incinerators. If practical, the advantages may be:

a) Avoid purchase of new or separate capital equipment.
b) Avoid fuel and operating costs of separate afterburner.
c) Recover heating value of fuel portion of fume.

Whether such a system is practical will depend on satisfactory resolution of the following problems:

a) It may not be satisfactory for the fume source unit to be made dependent on the continuous operating of the heater or furnace. Possibly the fume stream might be vented temporarily if the furnace should be shut down. (This may not be a problem if two or more furnace or boiler units are in operation, so that the fume stream can be shifted from one to the other if required.)

b) Fuel value or oxygen addition rates brought in with the fume should represent a small fraction of that regularly consumed in the furnace; otherwise, rather elaborate interlocking and regulating controls may be required. Such controls might include continuous oxygen and unburned hydrocarbon analyzers, which can be very expensive and have questionable reliability, in order to maintain the fuel/air ratio within safe and operable limits. (The problem here arises from the fact that ordinarily the air supply is made slave to the fuel supply, which is controlled by the load demand on the furnace or boiler. The addition of the fume may add an appreciable "secondary air" supply without being subject to the control action of the fuel regulator. The added fuel that may be present in the fume is not a major problem, since the temperature controller will cut back on the normal fuel supply to compensate.)

c) Fume constituents should not be such as to substantially alter flame patterns and flame radiation in the furnace. Sodium salts and highly unsaturated aromatics can make a flame more luminous, which might lead to excessive local heat transfer rates and burnouts.

d) Furnace brickwork can tolerate no fluxing elements, such as from suspended salts of sodium, potassium, vanadium, etc., if the brickwork is held at a temperature above the melting point of the salt materials. The molten flux dissolves away the brickwork; it also attacks the oxide film that protects metal against oxidation.

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e) Furnace structures, boiler tubes, etc., will be vulnerable to attack from halogenated hydrocarbons and sulfur-containing fumes. Metal parts are usually of low-carbon steel, may be expensive, and require considerable furnace outage time to replace.

f) Plume-forming pollutants should be avoided, such as streams containing phosphorous, or high in sodium and potassium.

g) The furnace must provide good fume destruction. By-passing would likely be a problem with a large unbaffled furnace. Also, the temperature may drop below 1500°F in some process heaters. The size of the furnace must be sufficient that the addition of the fume stream, when treated as a secondary air supply, does not cause the temperature of the mixed gases in the combustion chamber to fall below the required fume destruction temperature, even when the furnace is being operated under its minimum load condition.

h) The presence of fouling materials that could deposit on the tube walls should be considered. Fouling of these surfaces would reduce the capacity of the furnace or boiler by interfering with the heat transfer, and frequent maintenance would be necessary.

A full resolution of all of these problems should be apparent before use of this approach to incineration should be considered as a workable alternative. Reference 5-1 states that boilers have been used as afterburners successfully to control visible emissions from meat smokehouses and odors from rendering cookers; they have been used in refinery processes involving cresylic acids, hydrogen sulfide, mercaptans, sour water strippers, ammonia compounds, regeneration air from treating plants, oil mists and vapors from process columns. It seems unlikely that a large number of occasions will arise where the right combination of fume sources and suitable furnaces or boilers can be found.

5.2 Manner of Introducing the Fume Stream Into the Furnace

The variety of fume streams, and the furnaces, boilers and associated burner combinations is so large that no detailed recommendations can be made. Presented here are a number of suggestions and comments. They apply to handling fume streams which contain less than 25% of the LEL fuel value; streams containing more than this amount of fuel should be treated as fuels and handled in separate "waste heat boilers," or flared, with full regard for the special problems of handling fuel and air mixtures that might be within the combustible or explosive range.

5.2.1 Furnaces Using Premix or Diffusion-Mixed Burners (Gas or Oil Fuel) With Natural Draft, or Induced Draft Blowers

Commonly used in oil refineries and chemical plants, this type furnace offers the best expectation for incinerating a fume stream. If the combustion air is induced through the burner, however, the substitution of a fume stream for the air supply is possible only if it is non-corrosive, non-fouling, and if its oxygen content is very nearly the same as ordinary air. If the furnace uses a secondary air register, separate from the burner, it should be introduced through a tee into an extension of the register, as shown in Figure 5-1 (Reference 5-1), thereby taking advantage of the existing arrangements for insuring that the fume will be well mixed into the combustion gases. If there is no secondary air register, then one must in effect be created. It should be located in the immediate vicinity of the burners; Figure 5-2 suggests good and poor locations. The bottom entry immediately below the burner shown is suggested for use where fouling conditions are encountered (Reference 5-1), however, this would not be so if the furnace floor were located at ground level, since the ductwork would not be accessible for cleaning; also it would not serve well if condensible material is present, since a sub-floor channel would serve as a liquid trap.

Care must be taken if fume is introduced as additional secondary air to insure that adequate draft capacity is available to handle the increased flow of flue gases. Fumes containing about the same amount of oxygen as fresh air can merely displace the normal secondary air, with adjustments to the secondary air dampers; in this case no added draft capacity is needed. However, if the fume stream is deficient in air, the same amount of oxygen will be required in the furnace (in fact, a little more, since the furnace efficiency will be reduced by the extra gases that must be heated to the flue temperature), and the extra blower or draft capacity must be supplied. If large amounts of inert gas (CO_2, N_2, H_2O (vapor)) are added via the fume, burner stability may be reduced.

A reformer furnace, although commonly using premix gas burners, would be a poor candidate for incinerating fumes. This type of furnace uses a large number of burners which are individually adjusted in order to achieve a close control of the heating distribution to the tubes in the furnace. Possibly the fume might be introduced into a plenum which supplied air to all of the burners, if such a plenum could be applied, but this would interfere with access for lighting-off and adjusting the individual burners, as is common practice.

5.2.2 Furnaces With Burners Using Forced Draft Blowers

The problem here is that most fumes are available only at or near ambient pressure, and they would have to have their own blower or else pass through the furnace blower in order to enter the furnace. Blower units are particularly subject to difficulty in fouling services, and this arrangement is not recommended if either fouling or corrosive conditions are present. If the fume is non-fouling, an arrangement such as in Figure 5-3 may be considered, but such an arrangement is not easily applied when the burner unit comes as an integral package that includes its own blower.

A steam-jet ejector may be considered as an alternate to a rotating blower. Properly selected and applied, it is less susceptible to fouling.

In all cases involving extensive changes to the burners, the control system, or the provisions for supplying draft, the manufacturer of the equipment should be consulted. He may have had previous experience with a similar problem, and be able to supply equipment modifications and recommend alternative arrangements.

He should be made a responsible party for any changes that effect the safety or functioning of the equipment he has supplied.

5.3 Hazards on Shutdown

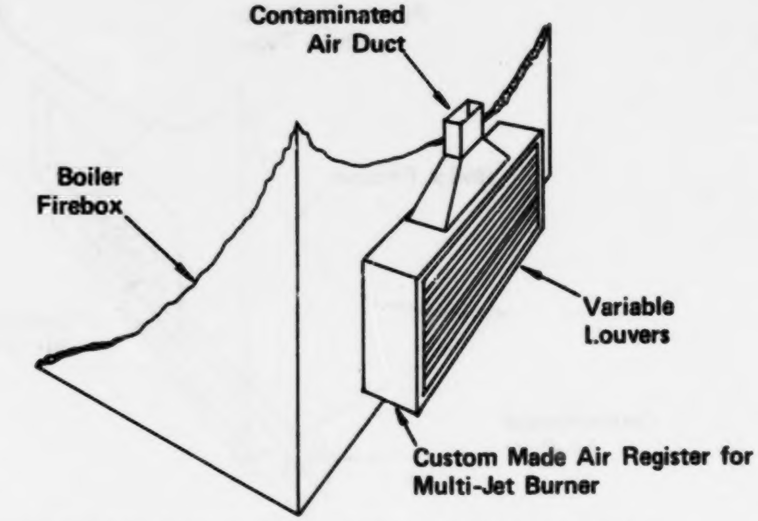
"Contaminants in most exhaust gas streams are normally well below explosive concentrations. In a few processes, however, combustible gas concentrations can accumulate during shutdowns with resultant explosion hazards on lightoff of the boiler. For instance, a batch of raw or partially cooked animal matter might be left overnight in a rendering cooker ducted to a boiler-incinerator. This could generate enough methane, hydrogen sulfide, and other organics to produce an explosive mixture in the ductwork leading to the boiler. If, subsequently, the burner were ignited without first purging the line, an explosion could occur. To avoid a rare possibility such as this, both the boiler firebox and the ductwork should be purged before igniting the burner.

"Some fire hazard is created by the accumulation of organic material in ductwork. Lines such as these must usually be washed periodically. The degree of organic accumulation can sometimes be reduced by frequent steam purging or by heating the ductwork to prevent condensation"(5-1).

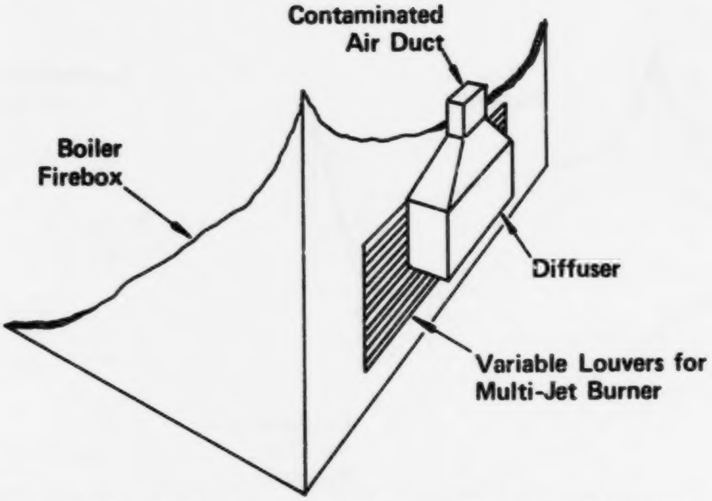
5.4 Verification Tests

When a process heater or boiler has been adapted to serve also as a fume incinerator, its optimum adjustment and satisfactory performance should be verified by tests both before and after the addition of the fume. The furnace may have been performing poorly before the fume addition; the fume stream will certainly not improve matters. Even if the furnace performs well for its ordinary function it may do poorly as an afterburner, as it is by no means easy to ensure the furnace will give the right combination of mixing, heating, and oxidation that will bring about the destruction of the fume. An improperly introduced stream may remain cool and bypass the main combustion chamber process, or it may be introduced in such a way that it partially quenches the flame before combustion is complete, forming CO and aldehydes. (See Chapter 14 for a discussion of the problems of mixing, and Chapter 10 for test procedures.)

Figures 5-1 through 5-3 follow


Contaminated Air Duct
Boiler Firebox
Variable Louvers
Custom Made Air Register for Multi-Jet Burner

A. Good method of introducing contaminated air to boiler firebox through a custom made air register. There is good flame contact. Contaminated air enters firebox through burner. Note: Type of burner is critical; contaminated air is portion of combustion air; not applicable where contaminated gases are corrosive.


Contaminated Air Duct
Boiler Firebox
Diffuser
Variable Louvers for Multi-Jet Burner

B. Poor method of introducing contaminated air from diffuser to boiler firebox through the burner air register. Diffuser restricts combustion air to burner. Moreover, louver may partially close restricting flow of contaminated air into boiler firebox.

Reference: LAAPCD Engineering Manual

Figure 5-1. INTRODUCING FUME STREAM VIA AIR REGISTER

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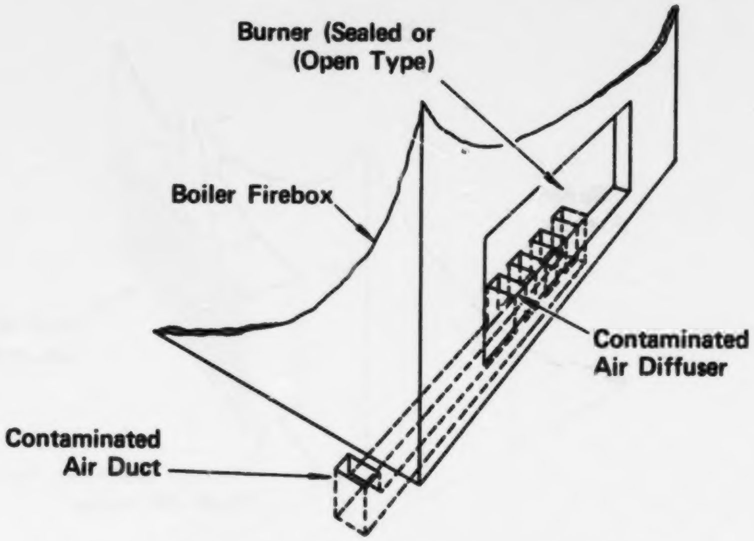
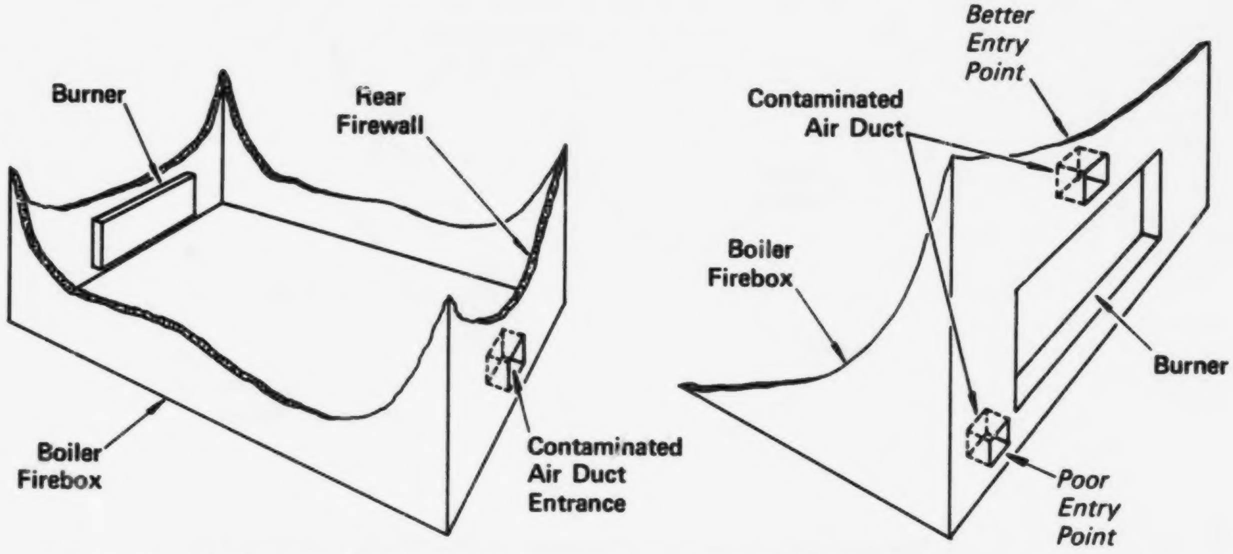

Burner (Sealed or
(Open Type)
Boiler Firebox
Contaminated
Air Diffuser
Contaminated
Air Duct

A. Boiler firebox showing entry of contaminated air through a diffuser in the floor near the burner. The possibility for avoiding bypassing is good. Note: Type of burner is not critical; contaminated air is secondary air for boiler; applicable where contaminated gases are corrosive.

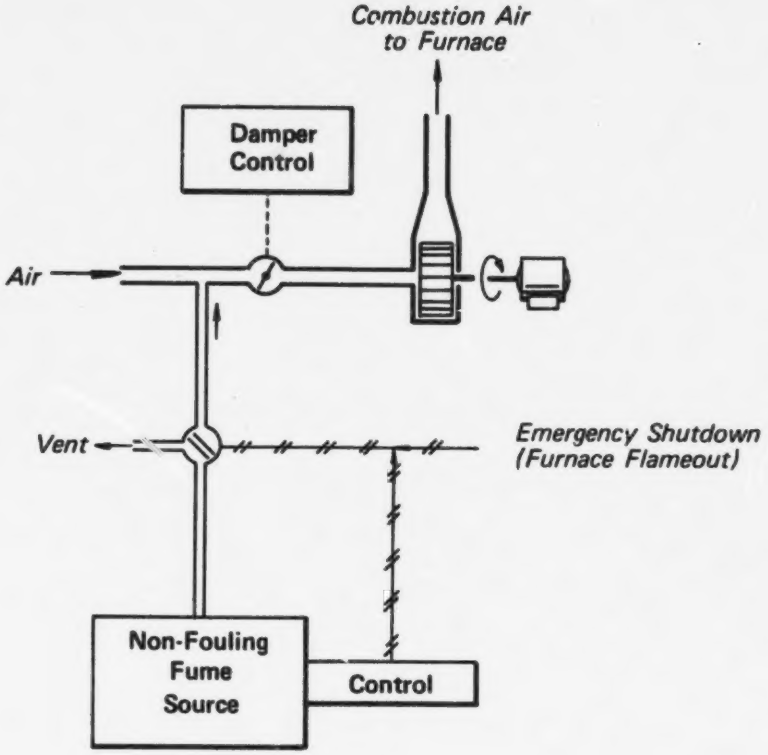

Burner
Rear
Firewall
Boiler
Firebox
Contaminated
Air Duct
Entrance
Better
Entry
Point
Contaminated
Air Duct
Boiler
Firebox
Burner
Poor
Entry
Point

B. Poor method showing entry of contaminated air near boiler firebox rear firewall. Mixing will be poor.

C. Boiler firebox showing entry of contaminated air through a duct at front of boiler.

Reference: LAAPCD Engineering Manual

Figure 5-2. DUCT ENTRY OF FUME STREAM INTO FIREBOX



Combustion Air
to Furnace
Damper
Control
Air
Vent
Emergency Shutdown
(Furnace Flameout)
Non-Fouling
Fume
Source
Control

Figure 5-3. SAFETY PROVISIONS FOR ADDING A FUME STREAM TO AN EXISTING FURNACE

Chapter 6. HEAT RECOVERY

6.1 General Considerations

This section describes methods and equipment designed to make use of the heat energy contained in the flue gases discharged from the afterburner, which would otherwise be wasted. A number of recovery methods are in use or have been proposed:

a) Heat exchange may take place between the hot flue gases and the cool fume stream, using a shell-and-tube (recuperative) or a rotary or cyclically regenerative heat exchanger, and thus preheating the fume stream before it enters the afterburner.

b) A portion of the hot stack gases may be returned, or "recycled" to the oven or process unit which is the source of the fumes.

c) The hot flue gas heat may be removed in some other manufacturing process unit, either directly (by exposure to the flue gases), by a heat exchanger to a circulating heat medium, such as oil, water, or molten salt, or to a steam generator for process steam, space heating, or power generation.

Since heating a large fume stream to a temperature as high as 1500°F represents a major consumption of fuel energy, and one that will continue indefinitely so long as the afterburner is operated, the economics and practicability of heat recovery should be weighed very carefully. Many factors must be considered:

a) Is it feasible? Dirty or foul fume streams may render heat recovery equipment inoperative in a very short time. Plant heat balance may not allow for an effective use of recovered heat.

b) Is it safe? On fume streams containing combustible material, excessive preheat by heat exchange with the flue gas might lead to flammable, or even explosive conditions. Recycle of hot incinerator stack gases could introduce contamination hazards in the process or oven equipment.

c) Is it dependable? If heat recovery equipment forces frequent shutdowns of the afterburner for cleaning and this in turn shuts down the plant unit, the economic value of any energy saving could be quickly nullified.

d) Is it economically justified? A proper economic evaluation should account for installed cost, operating costs (including fuel, operator and supervision, maintenance, etc.) interest charges, etc. Some guides for making this evaluation are given in Chapter 11. One manufacturer gives as rule of thumb that heat recovery equipment is economically justified only on afterburner units of more than 10,000 SCFM capacity.

e) Can the extra space and weight be accommodated? Heat exchangers are generally bulky, and so is the ductwork. If mounted on the roof, major structural changes might be required to carry the added weight.

f) Can deterioration of the heat recovery apparatus result in loss of control over the pollution source? Leaking exchangers can bypass fume material into the stack.

These questions, and others, are often complicated, and no simple rules can be made. All the conditions noted above must be answered affirmatively, one way or another, before a heat recovery system can be considered justifiable. The record of heat recovery systems is not very good; many have given trouble. The potential user should review proposed heat recovery systems with considerable skepticism. This is not to say he should reject the idea out of hand--rather that he should look for proven designs and applications which show successful use in a situation similar to his own.

A valuable aid for deciding whether or not to use waste heat recovery is to make a balance sheet of heat requirements and potential heat sources of the plant. Only if places can be found to use effectively the waste heat--in terms of load cycles, heating load quantities, and temperature levels--does it make sense to consider heat recovery schemes.

Table 6-1 summarizes the alternatives, the advantages, and limitations of these methods. The text material discusses the points of the table more fully.

6.2 Types of Recovery Equipment

6.2.1 Recuperative (Gas/Gas) Exchangers

6.2.1.1 Heat Transfer Effectiveness

The fume and the flue gas streams have about the same mass flow rates and specific heats, so they can be made to exchange heat with each other on nearly a degree-for-degree basis. Commonly this is done in a shell-and-tube (recuperative) exchanger, with one stream passing through the tubes, and the other stream passing over the outside. Usually the fume stream is put through the tubes, although manufacturers differ in their practice.

The performance of such an exchanger is measured by its "heat transfer effectiveness". The entering hot flue gas and the cold fume stream give the maximum temperature difference experienced in the exchanger, and this measures the maximum amount of heat that might be transferred. The temperature rise of the cold stream in the exchanger can never exceed this maximum difference, and the extent of its approach measures its heat transfer effectiveness, which is calculated thus:

$$E = \frac{Q}{Q_{max}} = \frac{C_p\ (T_{\text{fume stream leaving}} - T_{\text{fume stream entering exchanger}})}{C_p\ (T_{\text{flue gas entering}} - T_{\text{fume stream entering exchanger}})}$$

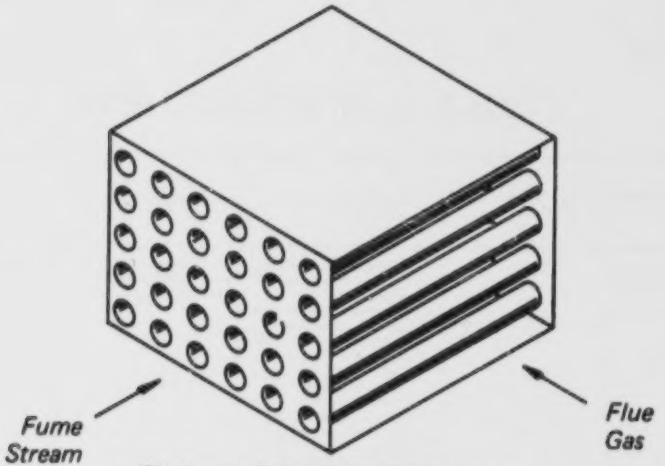
(C_p = the specific heat at constant pressure of the fume stream, Btu/(lb - °F))

Table 6-1 follows
Figure 6-1 follows

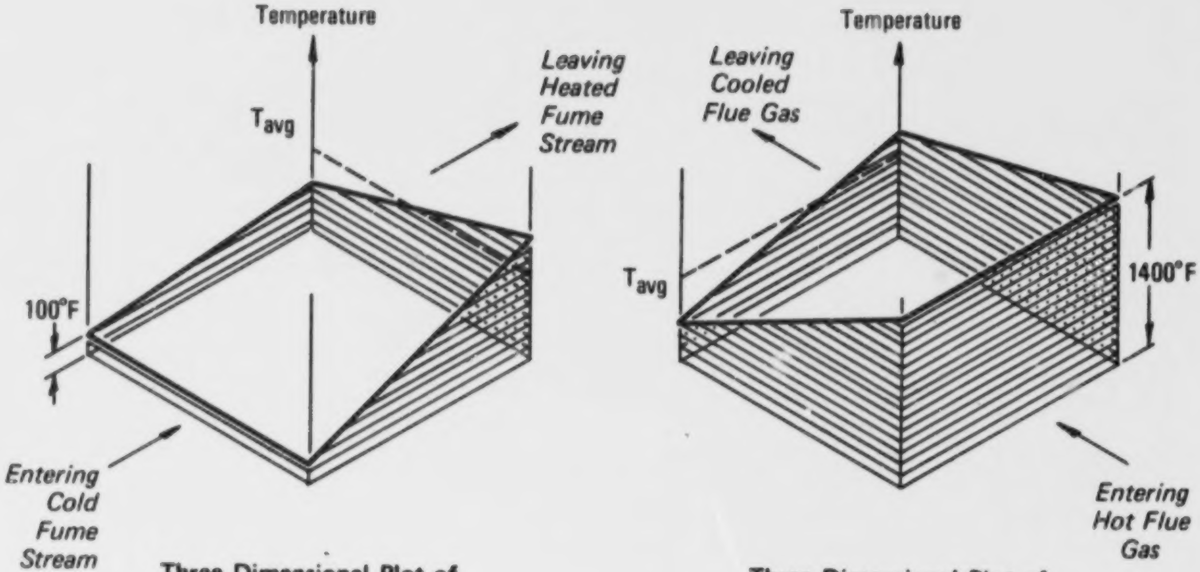
Table 6-1. FORMS OF HEAT RECOVERY

Type	Effectiveness Ratio, E	Additional Auxiliary Equipment	Limitations, Problems	Commonly Used For	Commonly Not Used For	No. Cases Reported in Survey Questionnaire Returned
Tubular exchanger Gas/gas	1 sta-50% max 2 sta-62% max 3 sta-85% max		1) May be easily fouled; frequent cleaning and maintenance. 2) Failures, differential thermal expansion. 3) Hot surfaces may crack or polymerize fume components, lay combustible deposit, initiate a fire. 4) Bulky, heavy, added roof load and/or floor space. 5) Corrosion if cools below dew point of flue gas.	Resin curing ovens Paint drying ovens Chemical plants	Rendering plants[a]	81 thermal 7 catalytic
Regenerative (rotary) exchanger	up to 85%		1) Easily fouled. Use only on relatively clean streams. 2) Burnout if failure on rotary drive motor. 3) Requires attention to pressure balance to control leakage at seals. 4) Avoid cooling flue gas to dew point, but otherwise is relatively insensitive to corrosion. 5) Ignition if overheat fuel-rich stream.	Odor control, water treatment units.	Any fumes containing oils, dusts, resins	2
Flue gas recycle to oven	~30%	Safety controls	1) Process must be compatible with flue gas (condensation? Sulfur in fuel? CO or CO_2? Reduced oxygen? Unburned fuel?). 2) Usefulness depends on temperature and heat requirements of fume generating process.	Resin, lacquer curing ovens, (if low solvent release) Litho ovens	Varnish cookers (combustibles hazard) ovens requiring human access. Smoke ovens (low temperature)	18
Steam, generators, boilers, water heaters	to 75%	Extra burners and controls; safety controls Extra ducting, blowers, controls	1) Ties steam generation to fume process and vice versa. 2) Match steam heating load to afterburning heat release. 3) Dew point and condensation on cold-water coils.	Plant steam supply CO burner and boiler (fluid bed catalytic cracking unit in oil refinery. Uses supplementary fuel firing) Plant heating and air conditioning units		(not tabulated)
Process heat--via circulating heat transfer salt (Hytec), oil, air	to 75%	Exchangers, piping, reservoir, pump, controls	1) Ties fume generation to process. 2) Matches process heat load to afterburner heat release (but can use supplementary firing for added process heat)	Asphalt blowing, preheating		(not tabulated)

[a] One manufacturer claims to have used heat exchangers successfully for rendering plants


Fume Stream
Flue Gas

Shell and Tube Type Exchanger
Cross-Flow Type


Temperature
T_{avg}
Leaving Heated Fume Stream
100°F
Entering Cold Fume Stream
Temperature
Leaving Cooled Flue Gas
T_{avg}
1400°F
Entering Hot Flue Gas

Three Dimensional Plot of Fume Stream Temperature

Three Dimensional Plot of Flue Gas Temperature

Figure 6-1. GAS TEMPERATURE PROFILES IN A CROSS FLOW HEAT EXCHANGER

90c

Clearly, the greatest heat economy will be achieved if E can be made to approach 1. However, this would be impossible without having an infinitely large exchanger. The effectiveness is also limited by the practical arrangements needed to confine and direct the gas flow.

6.2.1.2 Cross-Flow Exchanger

The need for a simple structure usually dictates using what is called a "cross-flow" arrangement, as shown in Figure 6-1. Although the exchanger can theoretically be made more compact (and heat transfer coefficients improved) by using small diameter tubes, the need for a low pressure drop and for easy cleaning usually dictates that all passages be an inch or so across, and tubes an inch, inch-and-a-half, or even two inches in diameter are common. In services that are subject to external fouling, tubes are arranged in vertical rows (not staggered) for easy cleaning.

Sometimes tubes with a rectangular cross-section and with fluted walls are used; the corrugations improve the heat transfer coefficient, and make a somewhat smaller exchanger. For a given overall exchanger effectiveness, there is little preference, one tube shape for the other, and the choice of surface is primarily a matter of cost. Tubes are anchored in a "tube sheet", and the tube bundle is enclosed in a metal "shell", which also confines one of the process streams. Hence, the term "shell-and-tube" exchanger.

From the standpoint of maximum heat recovery, the ideal arrangement for an afterburner would give counter-flow between hot and cold streams. Physically a true counter flow arrangement is difficult to obtain in a shell-and-tube construction (although a stacked-plate design approaches counter flow performance and is discussed later). The simplest and cheapest is the cross flow, and stacked or "multiple pass" cross flow units approach counterflow performance.

There are limitations to the effectiveness obtainable with a cross flow exchanger. Figure 6-1 shows that although the fume and flue gas streams each enter at uniform temperature, portions of the stream see different temperatures, and the leaving streams are not uniform. On one side the fume stream temperature may closely approach the flue gas temperature. On the other it sees an already cooled flue gas stream, and cannot reach as high a temperature. The average of all the streams may gain only a fraction of the maximum possible temperature rise (the effectiveness is less than unity). An effectiveness over 65% is possible on a single pass cross flow exchanger, as pictured on Figure 6-1, but only with much more surface area, and at the cost of a higher pressure drop; 40 to 50% is a commonly accepted compromise.

The exchanger can be arranged for multiple passes of either or both streams. A higher effectiveness is obtained if one of the streams is doubled back into a second pass, as in Figure 6-2, or even a third pass, with effectiveness values commonly running about 65% and 85%; the cost of the exchanger is roughly doubled or tripled to do this, and the blower pressure drop increased too.

Chapter 18 gives an approximate method for calculating the tube surface area required for an exchanger of specified effectiveness of the cross flow type. The size so determined should be used for preliminary evaluation only. Final judgment and equipment selection should be based on firm proposals or bids. Design proposals may be based on different assumptions as to fouling, pressure drop, heat transfer coefficient, etc., and on different materials from those assumed, and the cost quotation may differ appreciably from that estimated from the procedure given. The basis for assumptions on which bid proposals are to be made should be clearly defined if alternative proposals are to be judged fairly.

6.2.1.3 Stacked Plate (Counter Flow) Exchanger

A stacked plate exchanger may be encountered (Figure 6-3). Plates separate alternate channels for fume stream and flue gas, and a compact, near-counter-flow arrangement can be achieved. Heat transfer coefficients may be improved by embossing corrugations into the plates (thereby increasing turbulence and breaking up the insulating effect of the boundary layer at the wall surface). However, this arrangement suffers from severe local thermal stresses at the end sections, where much complicated welding is necessary to keep flue and fume streams separate. It is also difficult to clean. Its use should be limited to small temperature changes on clean service.

6.2.2 Rotary Regenerative Heat Exchanger

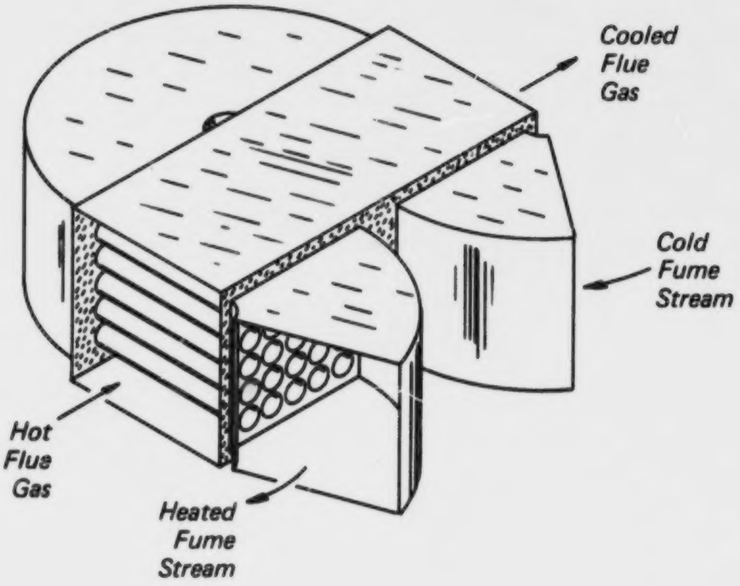
This type exchanger is diagrammed in Figure 6-4. A heat storage element made of corrugated sheet metal plates is mounted on a rotating cage structure and is passed alternately through the hot and cold streams. The plates are formed or stacked into a matrix structure, facing edgeways into the gas flow so that the gas may flow between them. The corrugations improve the heat transfer coefficient from gas to metal, and also serve as spacers to hold the plates apart. The matrix structure puts a large heat transfer area into a small volume, and an exchanger for a high heat transfer effectiveness for a large gas flow rate can be built into a very compact unit. Corrosion, so long as it does not impair the structural integrity or plug the flow channels, does not impair the performance.

This type of exchanger, when applied and operated correctly, can function very well. It is almost universally used for recovering heat from the boiler stack gases in central-station power generation units. However, there are important limitations and precautions to its use as an adjunct to the afterburner:

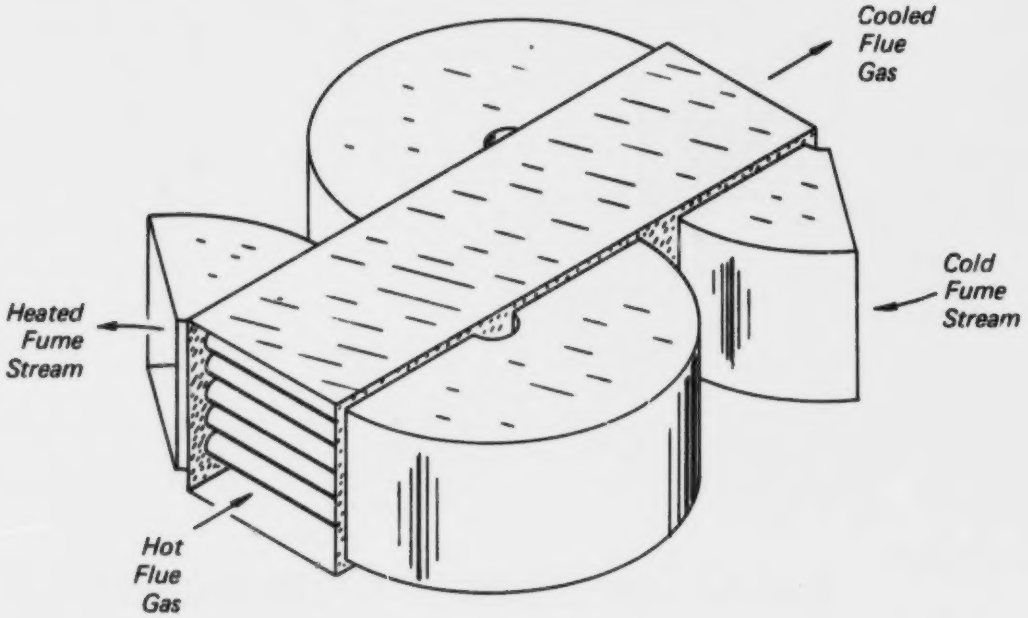
a) The small passageways between the plates are easily plugged. The fume stream must be free of condensible (and polymerizable) vapor, heavy smokes, or sticky particulates. Nevertheless, for some services, the passages are self cleaning (where non-sticky, dry particulates are encountered).

b) Isolating seals must be used to control leakage between hot and cold sections. These are subject to wear and warpage with time, and the leakage may become excessive. Effects can be minimized by arranging blowers so that pressure differences cause any leakage to flow into the fume stream where it will be recycled through the afterburner; this arrangement may be difficult to provide for some applications.

Figures 6-2 through 6-4 follow


Cooled
Flue
Gas
Cold
Fume
Stream
Hot
Flue
Gas
Heated
Fume
Stream

A. 2-Pass, Crossflow Exchangers (Arranged to Place Units Counter-Flow)


Cooled
Flue
Gas
Cold
Fume
Stream
Heated
Fume
Stream
Hot
Flue
Gas

B. 3-Pass, Crossflow Exchangers (Arranged to Place Units Counter-Flow)

Figure 6-2. MULTI-PASS COUNTER FLOW ARRANGEMENTS

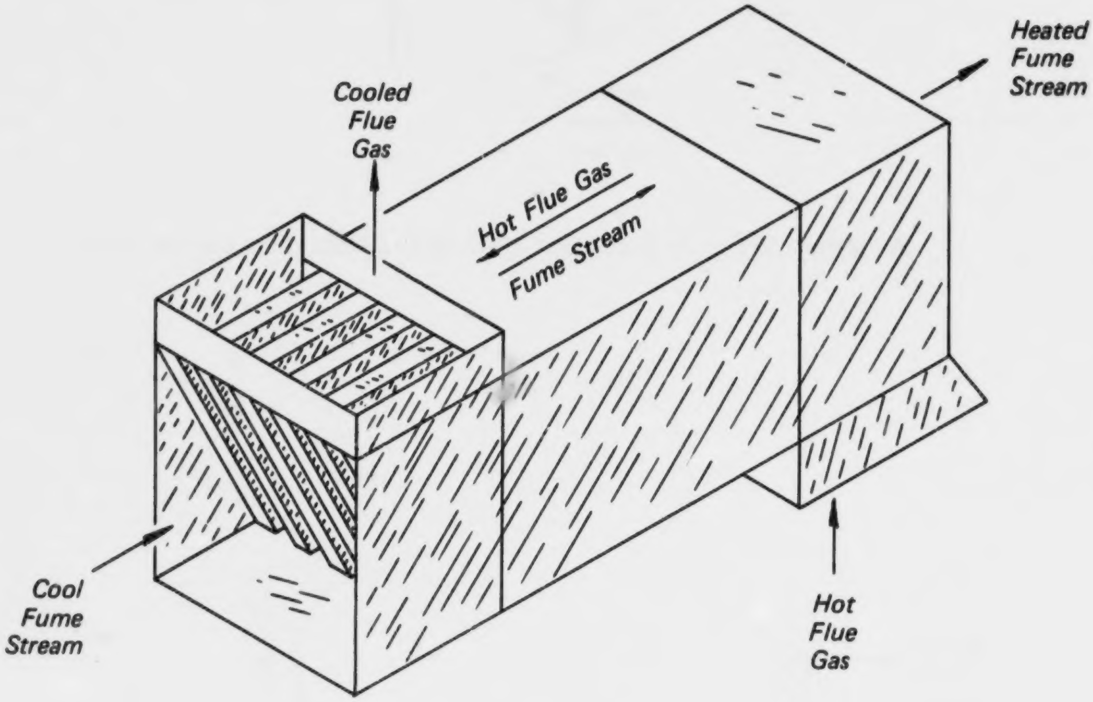

Heated
Fume
Stream
Cooled
Flue
Gas
Hot Flue Gas
Fume Stream
Cool
Fume
Stream
Hot
Flue
Gas

Figure 6-3. COUNTER-FLOW PLATE-TYPE EXCHANGER

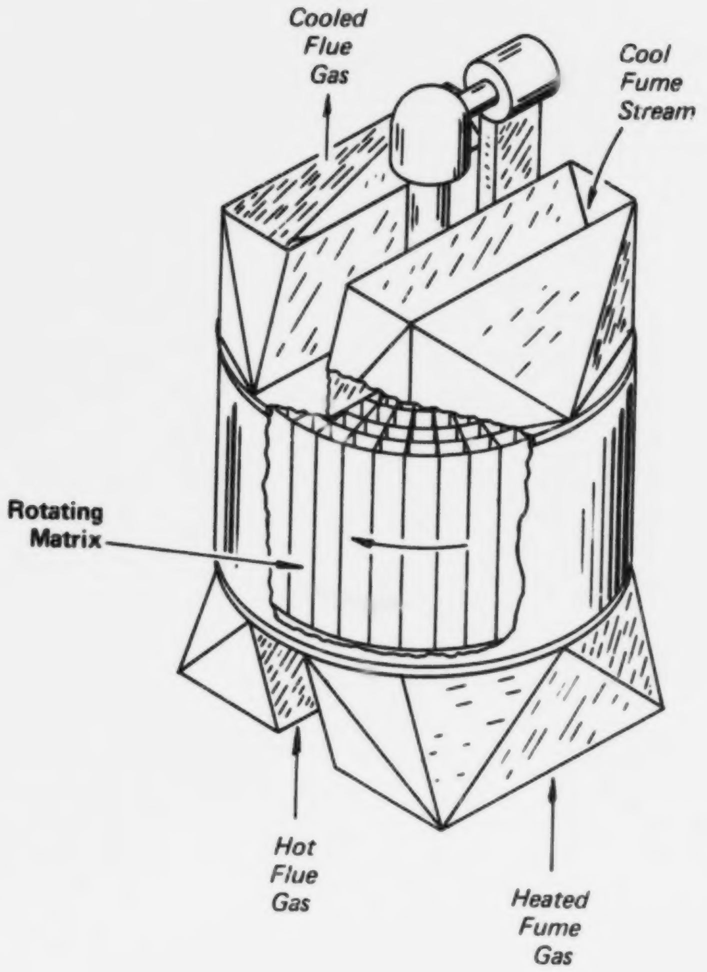

Cooled
Flue
Gas
Cool
Fume
Stream
Rotating
Matrix
Hot
Flue
Gas
Heated
Fume
Gas

Figure 6-4. ROTARY REGENERATIVE HEAT EXCHANGER

92d

c. As the matrix revolves, gas trapped within is carried from one flow channel to the other. The effect is minimized by blank areas in the cover plates. Ordinarily this trapped gas would bypass the afterburner and reduce its efficiency. As noted in (2) above, the flue gas pressure should be the higher, and the size and position of the blank area adjusted to give a positive sweep-leakage of flue gas into the fume gas side.

d. Since the seals must be close-fitting to avoid excessive leakage, they are sensitive to the strains caused by thermal expansion differences between hot and cold streams. Especially for exchangers taking flue gas directly from the afterburner (at around 1500°F), care must be exercised in designing to minimize effects of warpage within the exchanger structure, and of strains introduced by expansion of the attached ducting. Some early installations using the rotary regenerator have experienced mechanical difficulties on this score. Some manufacturers restrict use of this type exchanger to a second stage of heat recovery, with a recuperative type for the first stage; this exposes the structure to a smaller temperature difference and results in lower stresses and less warpage.

e. An inadvertant shutdown on the rotation of the matrix will allow it to be heated to stack gas temperatures. This could be damaging. Also, if any coating of combustible material has accumulated from the fumes, a fire might be started. A burner shutdown interlock system should be provided to protect against this type of failure, or else the rotor should be designed to withstand the maximum temperature of the flue gas.

f. Although corrosion effects are not as immediately serious as in a turbular exchanger, they should be guarded against. The metal of the matrix should never be allowed to operate below the dew point temperature at any point; preferably a 50°F margin above dew point temperature should be allowed when corrosive materials are present.

g. Because of the high effectiveness ratios to which this exchanger is best suited, the fume stream exit temperature becomes high, and combustible vapor streams that are well below the combustible limits at room temperature may ignite at the exit temperature. This is important for fume streams having a high fuel content.

With these limitations in mind, the rotary exchanger should be considered for installations on clean service, where high heat transfer effectiveness ratios are justified. Examples would be for odor control, or for bacteriological sterilization. A manufacturer who makes both the rotary regenerator and the shell-and-tube type exchangers reports that, as a rule-of-thumb, the rotary type is more economical to build and install when the desired exchanger effectiveness is over 60%.

The design of the rotary-regenerator type exchanger is complex, and should not be undertaken by the ordinary user. References on design and performance are:

Basic data on heat transfer and pressure drop of matrix: V. K. Migay, Heat Transfer - Soviet Research, Volume 2 No. 4 July, 1970 pages 151-156.

Rotary heat exchanger: W. M. Kays, Compact Heat Exchangers, McGraw-Hill, 1964. See pages 27-30, 44-49, 58-62, 70-71, and 259-260.

6.2.3 Cyclically-Operated Packed-Bed Exchangers

Closely related in principal to the rotary regenerative exchanger, although very different in physical arrangement, is the use of a pair of packed-beds, through which the cold and hot streams alternately pass in a switched cycle. The packed bed consists of screen-supported ceramic balls or saddles (see Perry's Handbook of Chemical Engineering for a more complete description of the packed bed and common packing materials). Normally used for catalytically controlled chemical reactions, the system may also be used as a heat exchanger with high effectiveness (how high depends on the thickness of the bed and the pressure drop available). One manufacturer offers the packed-bed exchanger for heat recovery on afterburners, and quotes design effectiveness ratios of 75% to 83%.

Advantages of the arrangement are:

a) Packing material is relatively inexpensive

b) Packing material is relatively inert, non-corroding

c) Exchanger bed is compact. There is a lot of exchange surface area and thermal storage capacity (mass x specific heat) in a small volume.

Disadvantages are hypothesized, since there are an insufficient number of installations to give a meaningful record in afterburner service;

a) Switching valves, some of which must work in a high temperature environment, are potential trouble sources.

b) During switching, the gas trapped in the bed from one stream is carried into the other. This would give a bypassing around the afterburner and a reduced system efficiency for controlling emissions. The manufacturer proposes to avoid this trouble by purging with clean air back into the fume stream. The arrangement has not been adequately demonstrated.

c) The packing, although not corroded, may be coated and insulated by fouling material carried in the fume stream, and the interstices blocked. Unless a pilot test can be arranged to prove otherwise, using the proposed foul stream, this exchanger should not be considered for use in a fouling type of service. Under some circumstances fouled material can be cleaned from the bed by burning.

Thermal performance of the cyclically operated packed bed exchanger is given in Compact Heat Exchangers (op. cit.) pages 27-30, 35-37, 44-49, 131. Pressure drop and heat transfer data on saddle packing should be obtained from the manufacturer, or from Perry's Handbook (op. cit.).

6.2.4 Recycled Flue Gas to Process Unit

This method of heat recovery has been used successfully on baking, curing, and drying ovens. It should be considered wherever large volumes of hot air or hot gas can be put to use in the process. Usually the oven temperature operates well below that of the afterburner stack, so that only a fraction of the stack gases can be recycled to the oven, being tempered on the way by mixing with fresh air to match the required oven temperature. The oxygen in the stack gas is partly depleted, hence the fume stream coming from the oven will also be reduced in oxygen content, and the burner flame stability should be verified. If a high recycle ratio is used (i.e., a high fraction of the flue gas is returned to the oven), external air will probably be required by the burner to satisfy combustion needs--with adverse effects on the fuel economy, especially when using a raw gas burner.

Figure 6-5 shows a possible arrangement with recycling for heat recovery where no appreciable fuel contribution comes to the afterburner from the process unit.

Use of this recycle arrangement would not be desirable if liquid fuels containing sulfur, or other corrosive or contaminating materials were to be encountered in the system. Recycling produces a concentration effect on contaminants as well as a depletion of the oxygen, and these effects may not be tolerable to processes going on in the oven (some lacquers may be very sensitive to contaminants, for example).

6.2.5 Waste Heat Recovery for Process Use

Where process heat is required, and neither recycling flue gas nor heat exchange with the fumes entering the afterburner is feasible, a number of possibilities remain.

a) The hot flue gases may be used directly in some other process unit. For example, they might be used as a heat source for evaporating and concentrating a wet slurry, or for a reboiler in a distillation column.

b) Hot clean air may be obtained for the oven to which the afterburner is attached using an air/flue gas exchanger. (Figure 6-6)

c) Process heat loads may require steam, which can be generated from heat in the afterburner exhaust (see Section 6.2.6).

d) Process heat loads may be best satisfied through an indirect transfer medium, such as hot oil, hot water, or a molten salt.

Whatever the process use, alternate bypass controls should be included so as to permit operating the afterburner as a heat source even though the fume source is not running, or to permit the afterburner to function on the fume stream when the other process unit is down. Stacks should be designed so as to withstand the temperature of the flue gases coming directly from the afterburner.

The amount of heat recoverable from the flue gas can be estimated with the aid of Figure 6-7, which is based on the assumption of normal flue gas composition derived from natural gas fuel. Some deviation from this will come about from different fuels and fumes containing combustible components.

6.2.6 Waste Heat Boiler

The exit gases from an afterburner are normally at a temperature of 1200°F to 1500°F, and considerable heat recovery can be obtained by passing these gases through a steam generating unit, or waste heat boiler. Note that this arrangement does not substitute the boiler furnace for the afterburner (such an alternative is discussed in Chapter 5); the boiler is considered only as a separate heat recovery device.

If desired, the waste heat boiler can also be equipped with separate fuel burners and controls; in this way, both afterburner and steam generator may be operated independently or in conjunction, without impairing the flexibility of either. Such an arrangement is shown in Figure 6-8. The choice of whether to fire the boiler, and how much firing to do, will depend upon the relative capacity demands on afterburner and on plant steam load. Generally the added complexity of supplementary firing adds appreciably to the cost, since the unfired version can be a relatively simple arrangement.

The variety of boiler designs useable for waste heat recovery is large. Some versions are of the locomotive type, with straight tubes fitted into tubesheets (up to 250 psig, 20 M lb/hr). Still larger boilers (perhaps >10 MM Btu, or >10 M lb steam/hr) will have the water contained in the tubes, with a steam drum for disengaging. Superheating coils may be included with a boiler of this type. Boilers of this size are available as package units, shipped complete with burners (if used) and auxiliaries, such as blowers, firing controls, feed pumps, etc.

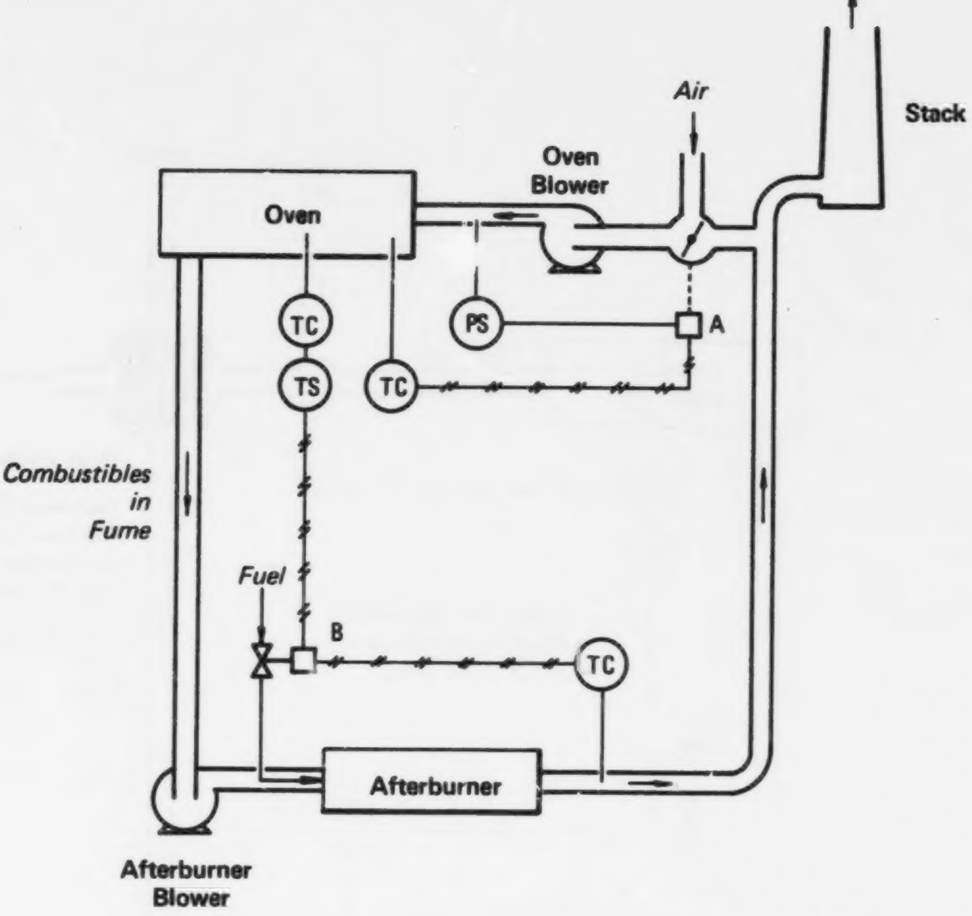
The amount of steam that can be obtained from the flue gases of an afterburner without supplementary firing is given approximately in Figure 6-9. This plot, like that of Figure 6-7, is based on natural gas fuel, without supplementary fuel value from the fume stream; it also assumes that the flue gases will not be cooled below 350°F, because of the possibility of condensing and corrosion on cold surfaces.

With the addition of supplementary firing, there is essentially no top limit on the size of the boiler. With whatever version is used, safety and controllability dictate using bypass around the boiler with blinds, so that either afterburner or boiler may be opened up for service. There should be a failsafe interlock control on firing that will provide emergency shutdown protection if there is a malfunction. Materials in the bypass/vent system should be able to withstand the afterburner exit temperature.

6.3 Fouling and Cleaning

A study of the equations of Chapter 18 make it apparent that heat exchanger surfaces must be clean for maximum effectiveness. A coating of

Figures 6-5 through 6-9 follow


Stack
Air
Oven Blower
Oven
TC
PS
A
TS
TC
Combustibles in Fume
Fuel
B
TC
Afterburner
Afterburner Blower

Notes:

A Two-Way Gate to Have Maximum Position Stop and Cut-Out Switch Actuated on Failure of Blower Pressure, to Avoid Overheating Oven.

B Emergency Override Cuts Fuel to Afterburner if Dangerous Over-Temperature in Oven.

A, B Should be Failsafe on Loss of Electrical Power or Instrument Air.

Figure 6-5. RECYCLED FLUE GAS TO PROCESS UNIT

96b

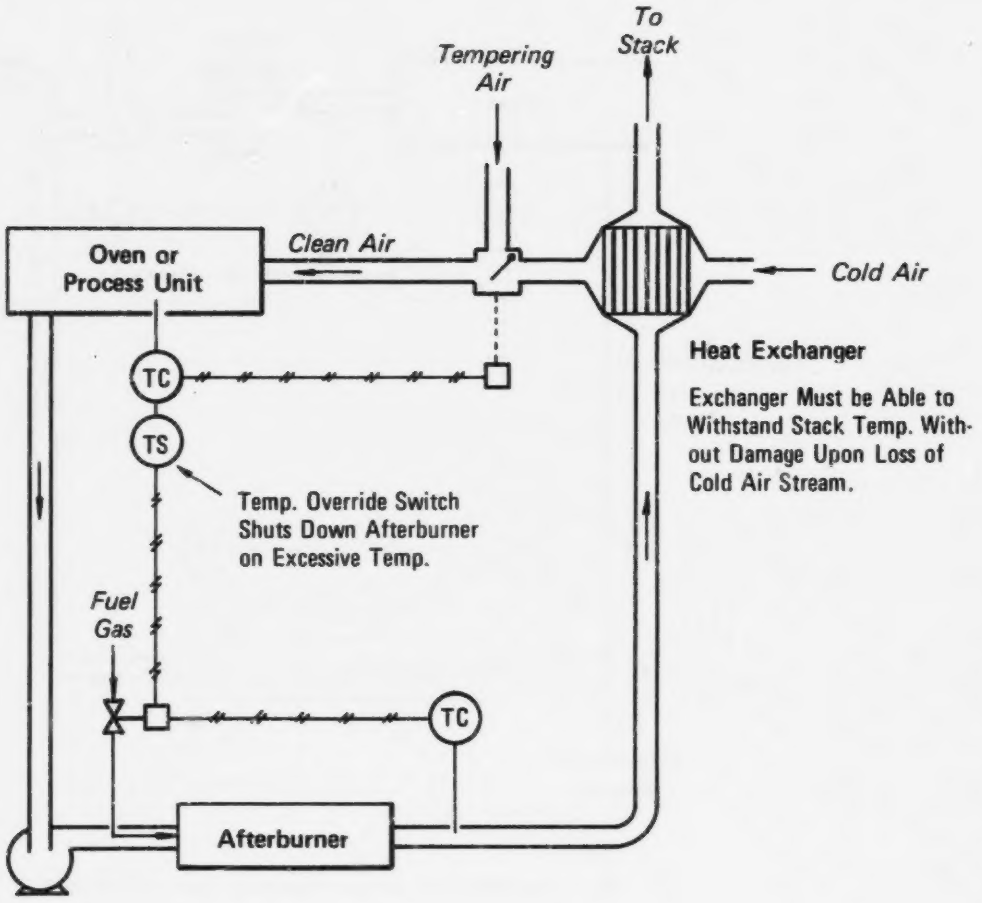

To Stack
Tempering Air
Clean Air
Oven or Process Unit
Cold Air
Heat Exchanger
Exchanger Must be Able to Withstand Stack Temp. Without Damage Upon Loss of Cold Air Stream.
TC
TS
Temp. Override Switch Shuts Down Afterburner on Excessive Temp.
Fuel Gas
TC
Afterburner

Figure 6-6. HEAT RECYCLE TO OVEN VIA HEAT EXCHANGER

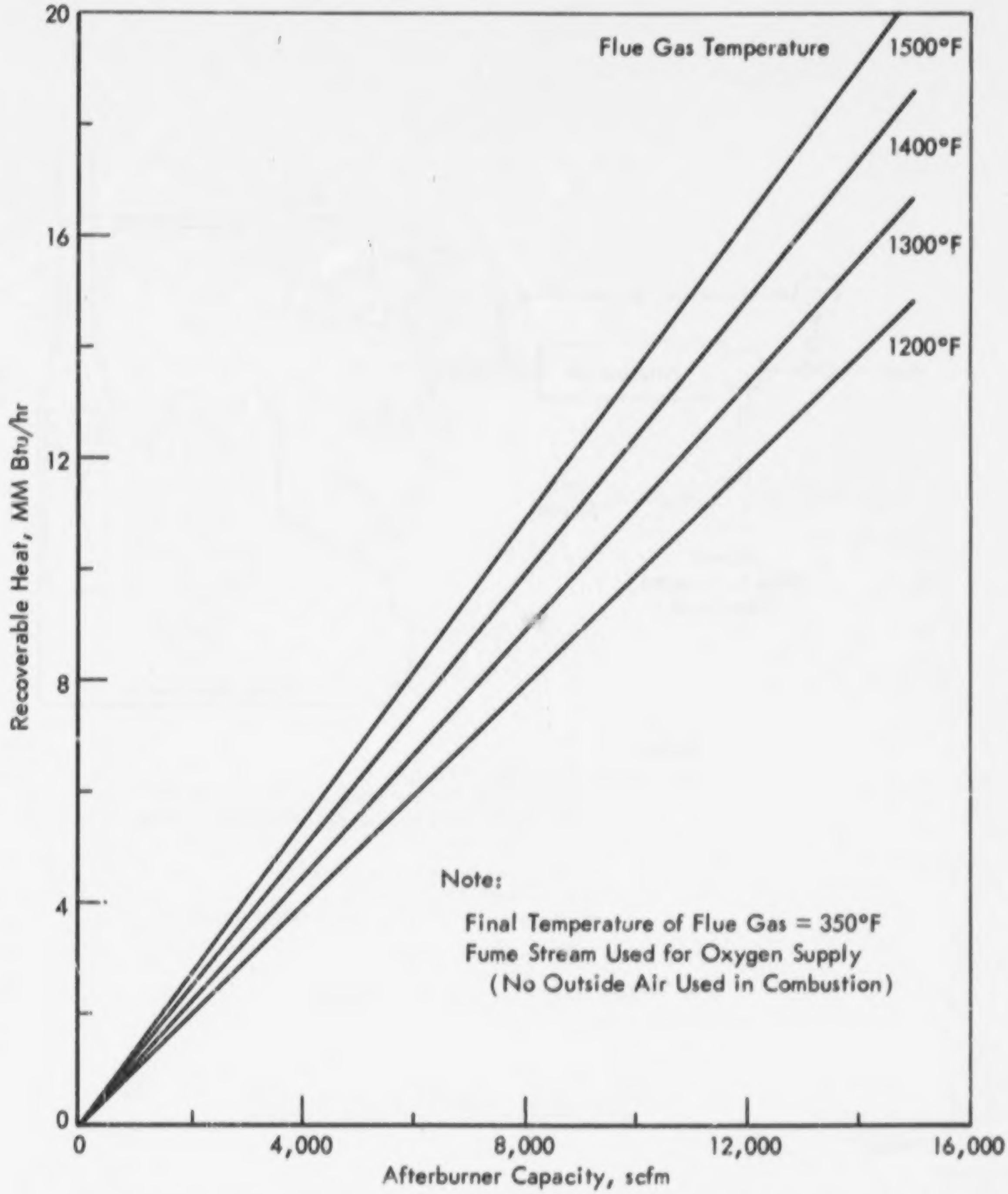
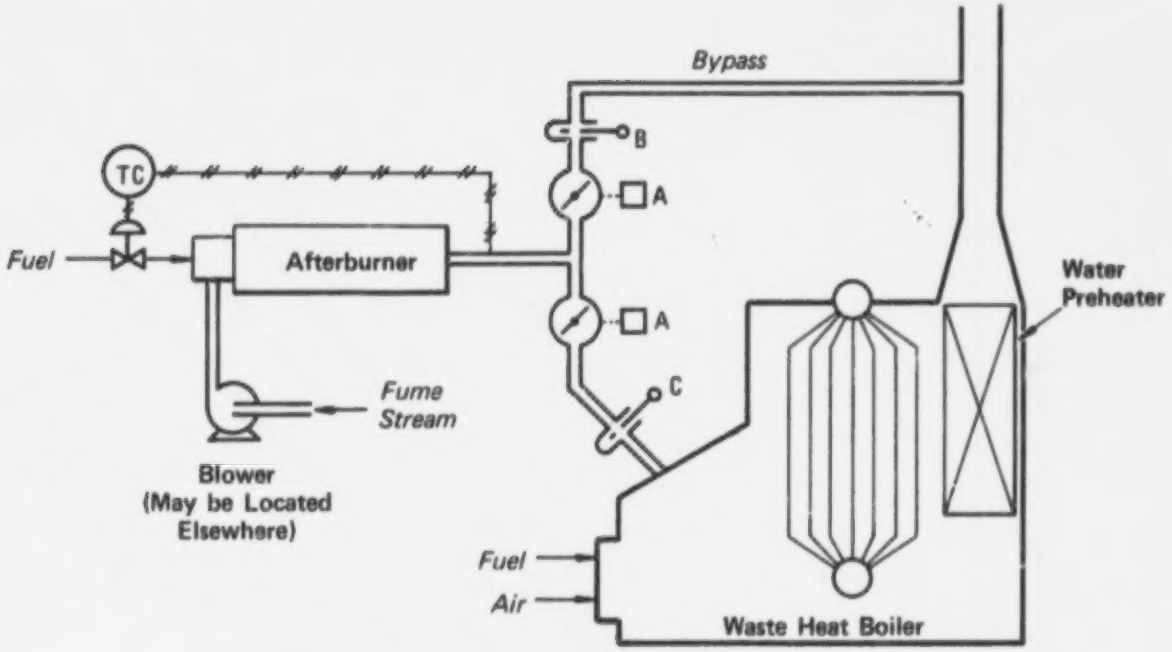


Figure 6-7. PROCESS HEAT RECOVERABLE FROM AFTERBURNER

96d


Bypass
B
TC
A
A
Fuel
Afterburner
Water Preheater
C
Fume Stream
Blower (May be Located Elsewhere)
Fuel
Air
Waste Heat Boiler

Notes:

A Automatic Diversion to Bypass

1) When Steam Pressure/Temp. Excessive and Boiler Fuel Cut to Minimum (Using Dual-Range Controller on Boiler Fuel).

2) Upon Loss of Boiler Feed Water.

3) Upon Loss of Boiler Flame.

4) (Closure of Both) Upon Shutdown of Afterburner.

C Blind When Boiler Not in Service.

B, C Blinds When Afterburner Not in Service

Figure 6-8. WASTE HEAT BOILER

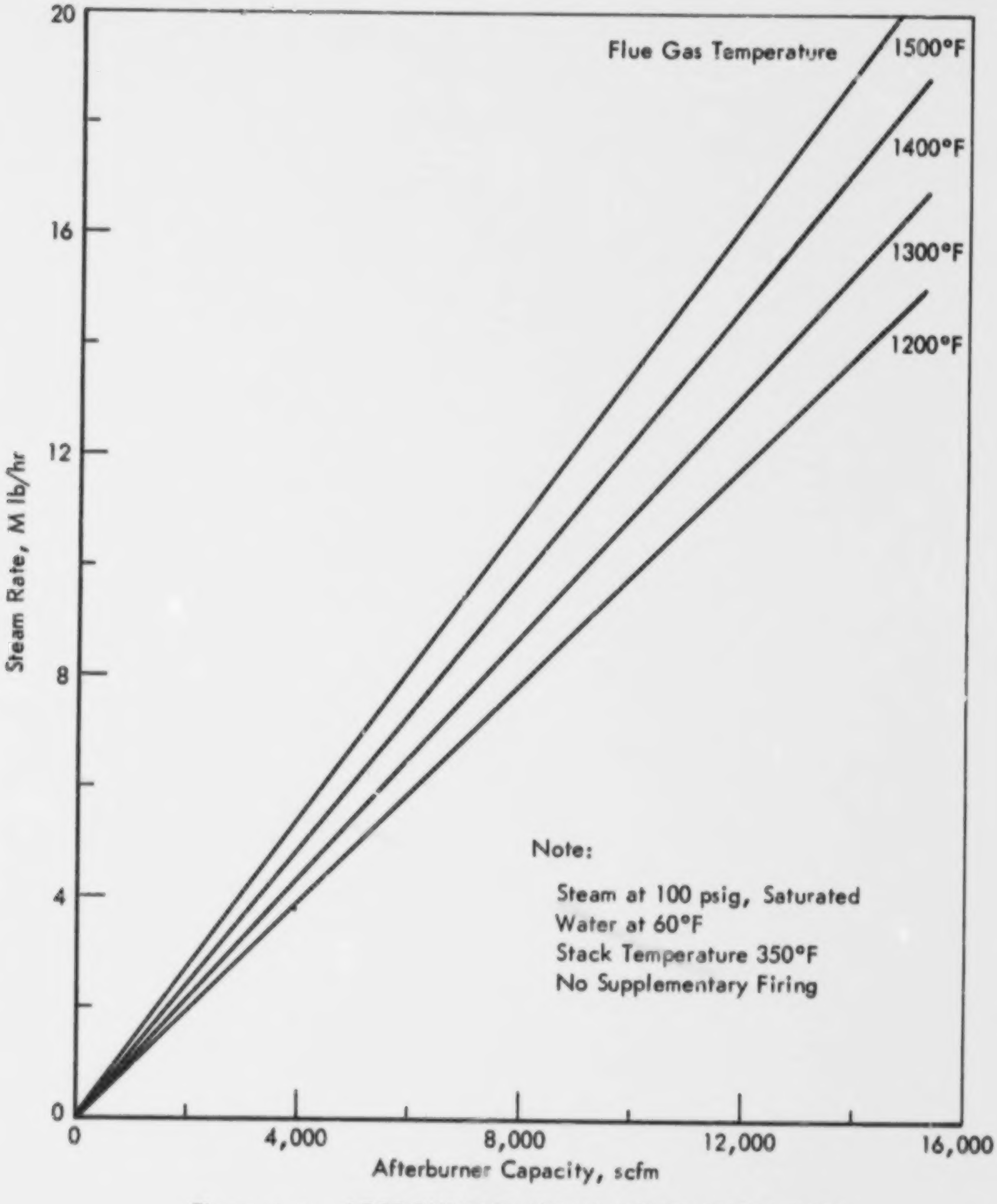


Figure 6-9. APPROXIMATE STEAM GENERATION RATE IN WASTE HEAT BOILER UNIT

96f

S-14121
67784

polymer, dust, carbon, on the exchange surface provides resistance to heat transfer, and less heat will be recovered. This is commonly accounted for in the equations by an empirical "fouling factor". An eighth-inch thick layer of polymeric, carbonaceous deposit on the fume side of a fume/flue-gas exchanger would give a fouling factor of about .12; a cross flow exchanger having an effectiveness of 50% when clean might have an effectiveness of only 42% when so fouled. The reduction in effectiveness, ΔE, represents a loss of fuel heat value, $\frac{\Delta E}{1-E}$, which can be a significant fraction for a high-effectiveness exchanger.

To minimize this loss, frequent cleaning may be needed. Most exchanger manufacturers prefer to put the fouling stream inside the tubes. Tough deposits may then be reamed out by a tool called a "turbine". If fouling deposits are less adherent, the fouling stream can be run over the outside of the tubes, and cleaning may be accomplished in place by water or solvent washing, steam lances, air jets, or by "rodding" between the tubes with hand tools. Rectangular tube arrays are preferred over triangular ones for this operation. The cleanliness after such an operation frequently leaves much to be desired, and placing the foul stream inside circular tubes is preferred. Cleaning by chemical or solvent treatment may not work in the afterburner, since the deposits are likely to be hard-baked to the point that they are not easily soluble.

The need for frequent cleaning also dictates that the exchanger and its ductwork be designed for easy access. This can be done with cover plates and access doors, or can be done by removing sections of ducting. There are no simple answers to accessibility for cleaning. The desire for low first cost and minimum weight and floor space favor closely coupling afterburner and exchangers; this often hampers accessibility for cleaning. It is expensive to bring in a hydraulic crane to lift ducting to one side. Flange joints are heavy and (especially at high temperature) develop leaks. The same is true of access doors. Using ample space for access complicates thermal expansion problems in the ductwork.

A user contemplating heat recovery for his afterburner when fouling conditions are expected would be wise to seek experience on similar installations on a closely related process in his industry. In the absence of such favorable experience, it is recommended that the prospective user view the use of a heat exchanger with a somewhat jaundiced eye, and include substantial allowances for frequent servicing in his economic evaluations.

6.4 Thermal Expansion and Thermal Stress

The metal temperature in a cross-flow exchanger varies from point to point (as viewed in plan-form in Figure 6-1), consequently, each tube row will have a different increase in length as it heats up to operating temperature. A rigid structure so heated will warp or tear itself apart (or, upon several cycles of heating, develop cracks from a "thermal fatigue" effect. Flexibility to accommodate this thermal expansion must be built into the tube fastenings in the tube plate (failure to do this in early versions of heat recovery equipment resulted in many examples of buckled tubes and ruptured tube fastenings. Sometimes tubes are attached to the tube plate with convolutions or "expansion bellows" to give the needed flexibility.

Attached ductwork should also be built to accommodate thermal expansion. This is especially true if any of the structure is to be built of stainless steel, which has an expansion coefficient nearly twice that of carbon steel. On large installations both afterburner and heat exchanger can be heavy, and duct connections must be provided with flexible expansion joints, or else one of the units (the heavier) should be rigidly tied at one point, and the rest of the equipment should ride on tracks that will allow free movement.

Where long hot ducts are used, as in the run from an oven to its afterburner, or in connecting the afterburner flue to a heat exchanger, the thermal expansion can be large. For example, a hundred foot long fume duct at 450°F, made of carbon steel, will expand about 3-1/2 inches over its cold length. A 25-foot long duct of stainless steel at 1500°F will expand about 5 inches. Provision must be made for this movement; otherwise an attempt to tie such a duct to an afterburner (which usually is built with a light gauge sheet metal shell) will surely cause a structural failure.

A fair degree of sophistication is required in the design of heat exchangers for use with large temperature differences. It requires a detailed knowledge of temperatures at each point in the system, and imagination in foreseeing the warping and distortion effects they may produce. Transient conditions during startup may impose severe thermal distortion effects (one part may heat up to operating temperature quickly, another part slowly). A major source of difficulty is that flow and heating distribution is not uniform in ducts; one side may run hot (from flame impingement?), the other side cool (from a recirculating eddy that sweeps a cold wall?). Flow through a heat exchanger bank may not be uniformly distributed, so that one group of tubes received more heat than anticipated, another less. Where possible a user would be wise to select a heat exchanger design that has proven itself under comparable service conditions.

Chapter 7 gives some data on thermal expansion effects and coefficients on some commonly used materials.

6.5 Double Wall Construction

This refers to the use of the afterburner shell as a heat exchanger element. The fume stream passes along and around the burner and holding chamber, between the outer shell and the inner ceramic or metal liner. Only a marginal reduction in fuel burned can be gained this way, although the cool fume stream helps make the outer wall less of a personnel hazard, and the insulation thickness needed is reduced. Some vapor condensation can be avoided, but there is a heavy penalty if the stream is a fouling one--the arrangement is very difficult (if at all possible) to clean. Also, it is a hazard, in that fouling deposits may accummulate, become ignited, and start a fire, releasing sufficient heat to damage or destroy the afterburner. Further, as discussed elsewhere, a 25% LEL fuel concentration in the cool fume stream may become a combustible mixture at elevated temperature; this mixture might be ignited by hot spots on the liner surface, and cause burning inside the wall of the afterburner. The prospective user should consider this design arrangement with skepticism.

Chapter 7. MATERIALS OF CONTRUCTION

The purpose of this chapter is to summarize the state of the art in selection of materials for afterburner systems, and to give data which will be useful in equipment selection and design.

7.1 Scope of Materials Review

Presented here is a brief review of materials commonly used in the construction of afterburners. No attempt has been made to review all the published literature on this subject. Only technical information readily available was used in this survey.

Some emphasis has been given to problems of corrosion, although most afterburner applications will give rise to no special problems of corrosive attack. Only a small minority of the afterburner installations involves the handling of corrosive fumes. The corrosivity can arise from the fuel used in the afterburner - that is, from burning oil fuels containing sulfur, or residual fuels containing ash-forming organo-metallics - or from the fume stream itself, which may contain sulfur, halides (chlorine, flourine, bromine, etc.), or very fine mists or particulate clouds of an endless variety of materials. Note that any appreciable quantity of some of these materials may require that the afterburner be followed by some form of scrubber; the problems of applying scrubbers or other collectors are outside the scope of this handbook. Because of the increasing scarcity of gas fuels and the need for using oil fuels and the increasing emphasis on the cleanup of chemical waste streams, some data on corrosion effects on metallic materials have been included.

Two classes of materials are discussed - ceramic and metallic. They are covered in two separate sections.

7.2 Refractories

7.2.1 Introduction

The normal configuration for construction of a thermal afterburner involves a steel outer shell, lined with a refractory material. The purpose of the refractory is to protect the steel shell from direct exposure to the effects of high temperatures and corrosive materials, and to improve thermal efficiency of the unit by limiting heat losses. Normal designs are based on the principle that the refractory serves as a thermal insulator, lowering the temperature from a maximum of more than 2000°F on the inside of the combustion chamber to a temperature of 250 - 400°F at the steel shell. At this temperature the shell retains most of its room temperature strength and is still hot enough to prevent condensation of water vapor on the surface, thereby lessening corrosive attack.

The refractory serves as a physical barrier to the high temperature combustion gases, allowing only minor amounts to percolate through to contact the steel shell. The gases are also cooled as they migrate through the refractory, so that corrosion by gaseous attack on the shell is greatly reduced.

Finally, the refractory serves to minimize thermal losses from the system, so that the burners may be run with relatively economical amounts of fuel.

The refractory may have any one of a number of chemical compositions and physical forms. Most refractories used in afterburners are made up of mixtures of aluminum oxide (usually called "alumina") and silicon dioxide ("silica"). For very special applications, bodies made up of magnesium oxide, chromium oxide, zirconium dioxide, or silicon carbide might also be used. The physical forms normally encountered are bricks, castables (refractory concrete), and dense boards made of pressed fibres.

Among the many factors to be considered in choosing materials for lining an afterburner are operating temperature, resistance to attack by combustion gases, thermal conductivity, material cost, installation cost, density, resistance to thermal shock, and vessel geometry. Additional operating factors, such as amount of operating supervision available for the unit, accessibility for repair, and frequency of shutdowns, also may affect materials selection. Each of these factors will be discussed here in order to demonstrate its importance in the materials selection process.

Refractory structures are heavy, with densities running 45 lb/cu ft for the light weight insulating firebrick and castable refractories up to 185 lb/cu ft for high alumina materials, and with wall thicknesses running typically from 4 to 8 inches (see Table 7-3). This weight adds considerably to the cost and difficulty of shipping, so that many afterburner manufacturers do not install the refractory until after the shell has been moved into place - especially on the large, custom-designed units. The installation of full masonry construction calls for a relatively skilled mason; an unskilled mason can make mistakes that will lead to premature failure. Once in place the added weight of the masonry can be an important consideration in the structural design of the building, especially if the afterburner is to be installed on a roof or elevated platform. Because of its light weight, fibre-block wall construction is being used in some afterburners (see Section 7.2.2.4). However, this material has limited strength, care must be used in attaching it to the metal walls, and gas velocities should be kept below 50 ft/sec to avoid the possibility of erosive damage.

7.2.2 Classification of Refractories

Refractories composed of mixtures of alumina and silica are by far the most common lining materials for thermal afterburners. They are generally classified in four groups, depending on their form during installation: brick, refractory concrete (castables), plastic refractories, and ceramic fibre materials.

7.2.2.1 Brick

Refractory brick consists of a mixture of clays, carefully selected by composition and particle size, which has been fired to a high temperature to drive off volatiles and to form strong ceramic bonds between the component particles. As a general rule, the maximum use temperature, thermal conductivity, and density all increase with increasing alumina content.

The maximum use temperature corresponds to the temperature at which a glassy silica phase within the brick softens, and the brick begins to flow. This temperature is indicated by reference to pyrometric cones. These cones are ceramic bodies of known softening point, which are placed in a furnace with the brick. The cone which slumps at the same temperature that the brick flows is then used as a reference indicator of maximum use temperature. Thus "cone 29" or "P.C.E. (pyrometric cone equivalent) 29" refers to a brick which softens in the same range as standard cone number 29, approximately 3050°F. Table 7-1 shows the classes of fireclay and high-alumina bricks, with corresponding compositions, cone ratings, and maximum use temperatures.

Table 7-1. CLASSIFICATION OF FIRECLAY AND HIGH-ALUMINA BRICK

Type	Composition (Wt % Al_2O_3)	Pyrometric Cone Equivalent	Maximum Rated Temperature, °F
Low-Duty	31 - 33	15 - 27	2600 - 3000
Medium-Duty	36	29	3050
High-Duty	38 - 42	31-1/2	3100
Super-Duty	41 - 46	33	3200
50% Al_2O_3	50	34	3250
60% Al_2O_3	60	35	3300
70% Al_2O_3	70	36	3350
80% Al_2O_3	80	-	3400
90% Al_2O_3	90	-	3500
Mullite	72	-	3350

A separate class of relatively low density brick, called "insulating fire brick", is used for its insulating properties, usually as a back-up for a denser brick lining. Brick is applied by the same hand lay-up techniques used in the building trades. The mortars used are special versions of the castable or plastic refractories discussed below.

7.2.2.2 Castable Refractories

Castable refractories are composed of mixtures of both raw and calcined clays of carefully chosen composition and particle size distribution. They also contain compounds, primarily calcium aluminate, which hydrate on addition of water. The castable behaves in much the same way as ordinary concrete, except that it forms very strong ceramic bonds after heating and dehydration. It may be applied by casting in forms, by gunning, or by troweling.

The classification of castables is usually on the basis of density and maximum use temperature; they may be obtained with ratings as high as 3200°F (even higher if exotic materials are to be used). See Table 7-3. There are light weight insulating castables, analogous to insulating fire brick, and denser castables used for more general applications. Special extra-dense abrasion-resistant castables are also available. These are particularly useful in systems where considerable particulate matter is entrained in flue gases. The temperature limit of a castable is usually stated in degrees, so that typical nomenclature involves terms like "lightweight 2700° castable."

7.2.2.3 Plastic Refractories

Plastic mixtures of ground fireclays with chemical binding agents are called plastic refractories. They are shipped wet from the manufacturer in sealed bags, and are usually applied by ramming with pneumatic hammers. They may be formulated to harden upon exposure to air ("air-set" varieties) or to harden when heated, by formation of a ceramic bond ("heat-set" varieties). The "air-set" types lose their strength on heating, and finally form ceramic bonds after exposure to high temperature. Plastic refractory temperature ratings are obtainable to 3200°F. Both castables and plastic refractories require special anchoring to the afterburner shell. Anchors may be either metal pins of various shapes welded to the shell, or refractory blocks attached to the shell by metal clips. Every refractory manufacturer and contractor has a particular preferred anchoring system, and as the choice is so dependent on the specific application, a contractor or manufacturer should choose the system to be used.

After firing at high temperatures, properties of bricks, castables, and plastics of similar composition are almost identical, except that the monolithic materials tend to be slightly less permeable than their brick analogs.

7.2.2.4 Ceramic Fibres

There has been a growing trend in recent years to use ceramic fibres for refractory linings. These fibres are usually of the kaolin composition (53%w SiO_2, 47% Al_2O_3) and are limited to applications below 2300°F. The fibres may be packed together to form a loose blanket, like mineral wool, or may be pressed with a sodium silicate binder to form a rigid board. The pressed materials have a somewhat lower (~1800°) temperature limit.

These materials have very limited strength, and must be installed with special clip fittings. They should not be used where maintenance calls for foot traffic. The gas velocities should be kept low (not over 50 ft/sec) in order to minimize the possibility of erosive damage.

7.2.3 Selection Criteria

7.2.3.1 Use Temperature

The single most important consideration in most refractory selection procedures is the ability of the material to withstand the design operating temperature. As shown in Table 7.1, most alumina-silica brick refractories have use temperatures far above the normal afterburner operating range. However, refractories near the burner may be exposed to temperatures in excess of 3000°F, so that special attention must be given to the refractories of material in this zone.

There is a common tendency to specify materials with a maximum use temperature far above the predicted operating conditions, to allow a considerable safety factor. This is not sound procedure, especially in the case of monolithic (castable and plastic) materials. The preparation of refractory brick usually involves firing to temperatures just below the maximum use temperature in order to develop ceramic bonds between individual grains. The process is not complete, and brick will continue to increase in bonding strength during use. The bond-formation process, known as sintering, is very sensitive to temperature, and use of a brick at temperatures well below its rated temperature will minimize any additional bond formation, leaving the brick relatively weak. Therefore, selection procedures should aim for materials with use limits close to design temperatures for the process.

In the case of monolithic lining materials, no ceramic bonding is present until the lining is thoroughly cured near a maximum use temperature. Because the cool face of the lining never reaches temperatures necessary for formation of ceramic bonds, strength is usually provided by the hydraulic bonds formed at low temperature. The center of a section of monolithic material usually has been heated sufficiently to destroy the hydraulic bonding but not enough to form ceramic bonds, and is relatively weak. If the hot face of the lining is never thoroughly cured by soaking at near maximum temperatures, the entire structure will be quite weak, and likely to spall.

The presence of a relatively weak center in a monolithic wall makes the structure a little more flexible than rigid brick construction, and often allows it to perform a little better in resisting thermal shock.

7.2.3.2 Resistance to Chemical Attack

Alumina-silica refractories are relatively inert, resisting chemical attack by a great many materials. As a general rule, problems of chemical attack are not a major factor in selection of afterburner materials. However, some components of flue gases can severely attack refractories, and any selection process must take these into account.

Most chemical attack causes refractory failure of three kinds. In one case, the presence of fluxing agents will cause melting of the refractory surface. If the melted material is removed by gravity flow or some other mechanism, then fresh surfaces are repeatedly presented for attack, and reaction rates can become so fast as to cause catastrophic failure. If the reaction product is not readily removed from the refractory face, further reaction may be essentially stopped. The reaction product will almost invariably have a sufficiently different coefficient of thermal expansion from the host refractory that severe stresses will build up during cool-down or subsequent start-up. The result is usually cracking and spalling (sloughing off of layers of refractory parallel to the hot face), with exposure of fresh surfaces to further attack.

The second primary mode of failure involves the formation of solid reaction products. In the case of refractories for incinerators and afterburners, operating temperatures are normally above the dew point for most corrosive species. Condensation can occur, however, in the pores of the refractory away from the hot face. The result of this condensation-reaction process is usually the formation of a solid reaction product with thermal expansion coefficient markedly different from that of the builk refractory. Depending on whether thermal expansion of the reaction product is greater or smaller, spalling will occur on start-up or shutdown, respectively.

A third mode of attack occurs only during shutdowns. If the reaction products of high temperature attack are hygroscopic, that is, attract water at room temperature, there may be swelling and spalling. Also, rapid heating on start up may cause evolution of steam from these compounds. When steam is evolved more rapidly than it can migrate out of the refractory, pressure builds up, and the refractory actually explodes. This same phenomenon, of course, occurs during initial curing of monolithic materials, and can be prevented only by very cautious initial heating rates.

The elements which are most destructive in attacking alumina-silica refractories are sodium, potassium, and vanadium. They react with aluminum oxide and silicon dioxide to form very low melting compounds. Sodium may also attack the alumina portion separately, forming the compound $Na_2O \cdot 11\ Al_2O_3$, which tends to promote bloating and spalling.

Calcium, zinc, phosphorus, iron, and cobalt are also often corrosive. These elements attack refractories mostly by formation of solid state (at normal afterburner temperatures) reaction products which later cause spalling during shutdown.

Specific operating experience with afterburners damaged by some of these elements includes:

In an automobile industry application burning used oil filters, the refractory was attacked by zinc volatilized from the galvanized cases. The attack was in the form of condensation behind the hot face, reaction to form zinc silicates, and spalling.

Sodium silicate binders used on wet-pack cardboard boxes were the source of sodium which severely fluxed afterburner refractories.

Crematoria and pathological incinerators, as well as some smokehouse afterburners, are often attacked by calcium and phosphorus compounds; the common reactions are bloating and spalling.

Phosphorus can be particularly troublesome in afterburner linings containing a significant fraction of iron oxide impurities. The common mode of attack is solid state reaction, followed by bloating.

In fibre wall constructions using relatively low density pressed fibre board, the presence of sodium in flue gases leads to embrittlement of the fibres at temperatures above about 1600°F.

7.2.3.3 Thermal Conductivity and Density

The thermal conductivity of refractories useful in afterburner linings varies over a wide range. The primary controlling factor is the density of the material. Table 7-2 lists some representative values for thermal conductivity of refractories.

Table 7-2. APPROXIMATE THERMAL CONDUCTIVITIES OF REFRACTORIES

Refractory Type	Conductivity at T, °F Btu/hr-ft-°F		
	200	1000	2000
Dense 90% Al_2O_3	25.7	19.3	17.7
Super-Duty Fireclay	8.7	9.3	10.0
High-Duty Fireclay	8.4	9.0	9.8
Insulating Firebrick (50 lb/cu ft)	1.8	2.4	3.2
Dense 60% Al_2O_3 Castable (3000°F rating)	4.6	5.1	5.7
Light Weight Insulating Castable (2000°F rating)	1.5	1.9	2.1

Insulating firebrick, which is often used as a back-up to a dense fireclay brick liner, is almost four times lower in thermal conductivity than superduty brick. Obviously, if no factors other than lining thickness and fuel economy were considered, dense brick would rarely be used. However, insulating firebrick offer much poorer resistance to chemical attack than do denser brick of the same composition. The low density of the insulating brick allows much greater penetration of corrosive materials, and the higher surface area increases reactivity.

Density, in addition to influencing thermal conductivity, and reactivity, is also of importance in economic considerations, for structural design, and in operating factors. The use of lightweight lining materials, which are usually cheaper on an area basis for equivalent thermal performance, also reduces shipping costs in those units in which refractory linings are shop-installed. The lighter linings permit the use of lighter steel construction, with attendant savings. In many afterburner applications roof mounting is desirable or even unavoidable; and the savings in lining weight permitted by lightweight materials, particularly fibre board construction, can make these applications economically feasible.

Typical densities of various refractory materials are shown in Table 7-3.

Table 7-3. DENSITIES OF REPRESENTATIVE AFTERBURNER REFRACTORIES

Material	Density, lb/cu ft
Super-Duty Firebrick	145
High-Duty Firebrick	135
Dense 90% Al_2O_3 Brick	180 - 185
2300° Insulating Firebrick	45
2000° Lightweight Insulating Castable	50 - 55
3000° Dense Castable	140
General Purpose 2200° Castable	120
3000° Plastic	140

7.2.3.4 Thermal Shock Resistance

Because of the brittle nature of refractory materials, they are subject to cracking when subjected to sharp temperature fluctuations. Because refractories are poor thermal conductors, the maintenance of strong temperature gradients is favored; and the stresses caused by differential thermal expansion from hot face to cold face can be considerable. Most manufacturers of refractory-lined equipment specify the maximum heating and cooling rates which the units can tolerate; the deciding factor in setting these limits is the resistance of the lining to thermal shock.

Thermal shock resistance increases with increasing thermal conductivity because stresses due to thermal gradients are lessened. Increase in density decreases the resistance to thermal shock, particularly among very dense materials. The presence of pores prevents spreading of microcracks; reduction of pore volume with increased density allows greater crack propagation.

The mineral mullite ($3\ Al_2O_3 \cdot 2\ SiO_2$) is an important material because of its excellent resistance to thermal shock. The combination of low thermal expansion coefficient, moderate conductivity, and high strength make this the best alumina-silica material for applications requiring good thermal shock resistance.

7.2.3.5 System Geometry

The shape of a reactor often strongly influences the choice of which class of refractory will be used for a lining. The use of monolithic materials permits design and installation of much more complicated linings than would be possible with brick construction. A large number of shapes are available in refractory brick, but often many of these are not shelf items, and long order times are needed.

As a first approximation, brick construction is limited to simple patterns - straight walls, cylinders, cones, and arches. Castable refractories may be used for any shape where installation of forms is possible. This is especially useful for providing baffles for combustion chamber mixing. Castables are particularly suited to small, relatively complex parts which may be fabricated in the shop. Plastic refractories can be used in any configuration where there is sufficient access to permit ramming.

The geometry of the reactor strongly affects the economics of the lining installation, as is discussed in Section 7.2.4 on costs.

7.2.3.6 Thermal Stress Loadings

The combination of ceramic and steel construction can give serious problems if proper allowances are not made for the relative thermal expansion of ceramic and metal parts exposed to temperature variations. This is discussed further in Section 5.3.6. Figure 7.1 shows the linear thermal expansion of common types of brick. Castable materials tend to exhibit less thermal expansion than brick but it isn't actually negative as often claimed. When castable ceramic cures it shrinks by an amount greater than subsequent thermal expansion when it is raised to 1400° - 1500°F. Therefore in afterburner applications castable refractory occupies the most volume when it is first installed and no thermal expansion need be allowed for.

7.2.3.7 Flame Impingement Problems

Ordinarily one would expect that flame impingement would not be a problem in afterburners provided the ceramic is selected so as to have a rating capable of withstanding the flame temperature. However, trouble could certainly be anticipated for situations where a 3000°F flame impinges a refractory that is rated at only 1500°F (as might easily be the case with some of the castable refractory materials). Moreover, there is a high probability for difficulty if the flame impingement gives an alternating exposure to the hot flame and the unmixed cold fume, brought about by the random and turbulent motion of the flame. The alternate heating and cooling of the surface will succeed in bringing about a slow spalling attritition. The problem is especially acute with oil flames because of their highly radiant character.

7.2.3.8 Installation Methods

The different types of linings call for vastly different installation methods; the choice of a lining is largely determined by the relative costs of the possible alternative installations.

Brick linings require experienced, skilled workmen for installation. Because of their relatively high weight, brick linings are almost never shipped installed, but are installed by refractory contractors at the field installation site.

Castable linings are much simpler to install, and withstand damage during shipment rather well. For this reason, plus their applicability to a wide range of geometries, and the economy and chemical resistance of lightweight castable linings, they are probably the most widely used materials in afterburners. Unskilled plant personnel can install (and repair) these linings, which are normally dried to about 200°F before shipping. This drying removes excess water and strengthens the hydraulic set of the lining.

Most plastic refractories do not have adequate strength to withstand shipping damage; and skilled installation is necessary. They offer few advantages over other lining types, and are relatively less common in afterburners.

Fibre linings offer the optimum properties consistent with plant installation and shipment: excellent mechanical shock resistance ease of installation, and light weight. They are nearly always shop installed.

7.2.4 Costs

It is almost impossible to place real values on costs of linings because of temporal and regional fluctuation in materials prices, freight rates, and labor costs. The values presented in the following sections are intended only as guides to indicate the range of costs that might be expected.

Figure 7-1 follows

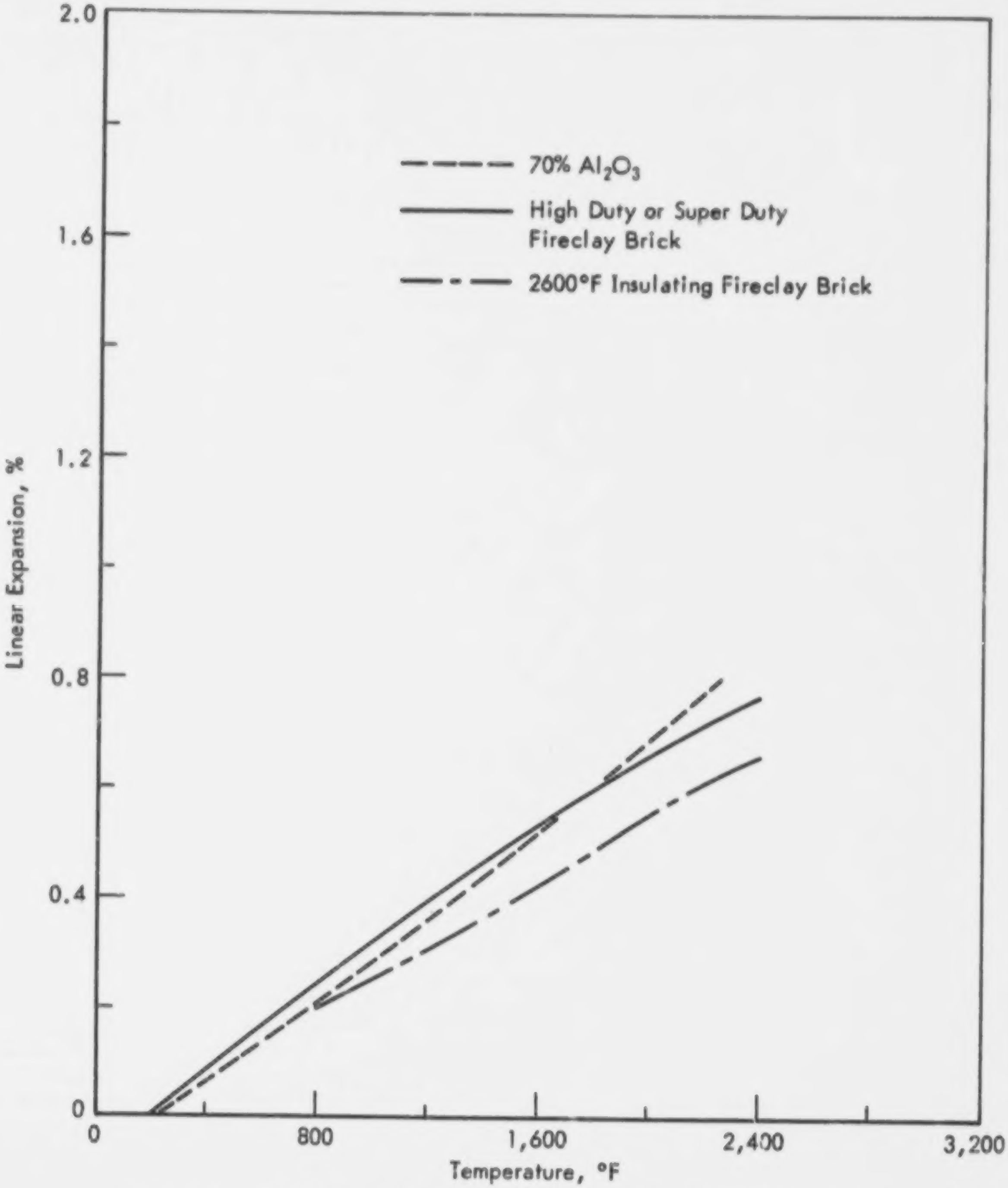


Figure 7-1. LINEAR THERMAL EXPANSION OF COMMON BRICK TYPES

108k

S-14121
67784

7.2.4.1 Materials

Monolithic materials may cost more than the equivalent brick on a thermal efficiency basis. That is, the amount of castable or plastic necessary to provide a given shell temperature may cost somewhat more than the amount of firebrick needed for the same job. However, for brick shapes other than very simple configurations, this advantage may vanish, and will also be offset by labor savings for monolithics. In late 1971, the following price ranges were realistic (all prices at point of manufacture): superduty firebrick, about \$250/1000 brick (brick prices for "9" straights," a brick 9" x 4-1/2" x 2-1/2"); 2500° castable, about \$150 per ton; and a comparable plastic about \$125/ton (wet). On a volume basis, these figures are equivalent to about \$4.50/cu ft for brick, \$3.00 to \$12.00/cu ft for castable (depending on density), and \$7.00 to \$10.00/cu ft for plastic.

7.2.4.2 Installation

The speed of installation of various linings is extremely dependent on geometry. For very simple configurations (walls, floors, etc.) manufacturers estimates range from 24 - 28 hours of labor/1000 bricks, and 10 - 24 hours/ton of castable or plastic. This would make installed costs of a simple superduty firebrick structure \$40 - 60/cu ft, and castables \$15 - 50/cu ft, depending on density; these prices do not include freight charges for the materials.

Fibre wall construction consisting of fibre blanket and mineral wool block back-up has been estimated in the trade journal at approximately \$15/cu ft, installed.

It should be stressed that labor for a complex installation of brick could force brick prices very much higher, and that the use of low-priced labor for castable installation could lower that estimate. Also, these figures do not take into account the differing thermal efficiencies of the materials. The real figure for comparison would be cost per square foot for a wall thick enough to provide the necessary shell temperature.

7.2.4.3 Maintenance

The maintenance costs of various lining types depend strongly on the mode of operation, amount of supervision, and accessibility for repair. In simple applications where the unit is rarely shut down, no corrosives are present, and reasonable care is exercised in heating and cooling rates, many years of trouble-free service can be expected.

If the unit is subjected to frequent over-temperature excursions, brick construction may have an advantage. When thermal shocking is unavoidable, the fibre approach may be very attractive.

It is probably somewhat more economical to make small repairs on monolithic linings than on brick; but so many factors, such as availablility of labor, accessibility, etc., enter into this situation that firm guidelines are dangerous.

7.2.5 References

There is very little published literature which bears specifically on the problems of refractory linings in afterburners. However, there is an enormous amount of information in background areas, such as calculation of wall thicknesses, or proper curing methods for monolithic linings. The references listed 7-1 - 7-6 are intended to provide convenient sources for detailed discussions of subjects only touched on here.

7.3 Metallic Materials

Ordinary low carbon steels are used exclusively for the structural components of afterburners. Many manufacturers use a gas-tight structural container to enclose a combustion chamber that is built of metal, but which is lined with ceramic material for heat insulation. For most afterburner services this container vessel is also of low carbon steel, as it experiences a temperature of no more than 400°F. Other units are designed with metal parts exposed to the inside of the combustion chamber; these units make extensive use of alloy steel construction. Units of this type do not expose the metal to the hot part of the flame; at most they require that the metal withstand the 1200 to 1400°F temperature of the residence chamber, and they have often been designed so that one side of the metal parts is cooled by the incoming cool fume stream, so that it experiences a temperature below that of the gas in the chamber.

The low temperature, non-corrosive applications that permit of carbon steel construction will not be discussed here, since they lie within the realm of common engineering experience.

The creep resistance and the creep rupture life of metal parts exposed to the high temperature levels becomes important, and data is given on some commonly used materials. Data is also provided on the oxidation resistance of metals at high temperature. Special problems of chemical environments likely to give corrosive attack are reviewed; also a related one of cool surfaces where corrosive conditions can arise because of condensation of components in the flue gas. Some thought is directed to the problems of dealing with thermal fatigue.

Reactions of metals and nonmetallic materials with corrosives is a highly complex subject. In this summary it is possible to give only preliminary consideration to the many combinations of corrosives and materials of construction that may be encountered. The purpose here is to indicate the factors that influence selection of construction materials. Any specific environment of interest would need to be considered in greater detail. Thus, the information presented is not to be considered as complete for final selection of materials for a specific afterburner application.

The afterburner may be called upon to handle vapors of varying composition. The combustion gases may contain a number of acidic components in addition to the gases commonly found in boiler or furnace operation. Thus, selection of materials of construction cannot be based solely on previous experience in other types of furnaces burning conventional fuels. The limitations on materials of construction that relate to the fire side of steam boiler environments will apply; but in addition, the environment may be more severe producing additional limitations on materials of construction.

The additional environmental components to be considered here include low concentrations of chlorine, hydrogen chloride, oxides of nitrogen, phosphorous pentoxide, sulfur compounds plus various metal oxides and salts such as Al_2O_3, Na_2O, NaCl, SiO_2 and oxides of potassium and cobalt.

7.3.1 Mechanical and Metallurgical Considerations at High Temperature

Metal parts within the heated enclosure must be designed to withstand the operating temperature. The effect of temperature on creep strength and stress rupture strength of a number of metallic materials of construction is illustrated by the data in Table 7-4 and Figure 7-2 and 7-3. It will be seen that creep is a significant factor in use of many metals at temperatures as low as 1400°F.

In addition to changes in creep strength and other mechanical properties, many alloys are subject to changes in metallurgical properties at elevated temperature. Specific comments on use of various common materials of construction at elevated temperature are presented below.

7.3.1.1 Carbon Steel and 1/2% Mo Steel

After prolonged exposure at temperatures in the 800° - 1300°F range, carbon steel and 1/2% Mo steel in general experience spheroidization of carbides and graphitization. Significant reductions in strength accompany both spheroidization and graphitization. Although spheroidization will affect plate steel uniformly, graphitization occurs preferentially in weld heat affected zones. Following elevated temperature service, e.g., delayed coking units operating in the 900°F range, 1/2% Mo steel has been found to be more susceptible than carbon steel to cracking. The minimum commercial alloy necessary to prevent graphitization is 1% Cr - 1/2% Mo steel.

7.3.1.2 Low Alloy Steels (1% Cr, 1/2% Mo to 5% Cr, 1/2% Mo)

These steels exhibit the normal reduction in tensile properties with increasing service temperature. However, if heated above the transformation temperature (on the order of 1300°F) and subsequently cooled rapidly, they have reduced ductility at lower temperatures. To prevent this decrease in ductility when overheated, the material must be cooled slowly (50 to 100°F per hour) down to 1100°F.

7.3.1.3 Aluminized Steel

Experiences with the use of aluminized steel in afterburner service have not been good. The aluminizing process is a coating process by which aluminum is applied to and diffuses into carbon steel or alloy steel as a surface layer.

The coating has good resistance to high temperature oxidation and to high temperature sulfur corrosion. Damage to the coating is likely to occur and in such cases metal areas that are not aluminized are also exposed to the corrosive environment. Aluminizing is not recommended as a substitute for the proper alloy to provide corrosion resistance. Its use may be considered if the coating is applied to increase the life of a component that can perform satisfactorily in the uncoated state. However, the aluminizing process may produce adverse effects on the mechanical properties of the base metal.

7.3.1.4 Chromium Stainless Steels (Types 405 and 410)

The 12% Cr stainless steels (Types 405 and 410) are generally satisfactory materials for use above 800°F. However, higher chromium steels - 17% Cr or more - are not acceptable because they are susceptible to "885°F embrittlement." This aging phenomenon produces extensive decreases in toughness and ductility in the high chromium alloys after long term exposure in the 700 - 900°F range. Because the minimum chromium concentration necessary to cause 885°F embrittlement has never been conclusively established, Type 410 stainless steel (13.5% Cr maximum) is to be preferred over Type 405 stainless steel (14.5% Cr maximum).

Close attention must be directed toward welding of the 12% Cr steels because variations in weld deposit composition can exert marked influence in mechanical properties, especially toughness. We are aware of no experience with large vessel fabrication involving solid 12% Cr materials. However, these steels have been used extensively as vessel linings.

7.3.1.5 Chromium Nickel Stainless Steels (Types 304, 321, 347 and Cast Stainless Steels)

Metallurgical factors associated with long term exposure at high temperature can result in greatly reduced corrosion resistance on subsequent exposure in liquid corrosives at lower temperatures. Due to carbide precipitation (sensitization), all of these stainless steels will be less corrosion resistant after prolonged exposure above 800°F. Thus, it is desirable to ensure that they are not exposed to liquid acid corrosives of the environment, even at ambient temperature such as during shutdowns.

Exposure of these steels in the temperature range of 1000 to 1600°F can result in formation of "sigma phase." This can markedly reduce subsequent room temperature ductility. However, above 1000°F, presence of sigma phase may increase tensile strength with ductility being only slightly affected.

Type 347 stainless steel is subject to pronounced weld heat affected zone cracking and is not recommended for elevated temperature service.

There are a number of cast stainless steel compositions for high temperature service. Any specific alloy would have to be considered in detail before it is selected for high temperature service. The HT alloy, for example, is susceptible to hot shortness, an embrittlement phenomena producing cracking at elevated temperature. The preferred composition HK-40 (carbon content 0.37 to 0.45%)

Table 7-4 follows
Figures 7-2 and 7-3 follow

Table 7-4. CREEP STRENGTH

Metals	Stress for Creep Rate of 1% in 10,000 Hours, kpsi								Stress for Creep Rate of 1% in 100,000 Hours, kpsi							
	Temperature, °F								Temperature, °F							
	400	600	800	1000	1200	1400	1600	1800	400	600	800	1000	1200	1400	1600	1800
Iron Base Alloys																
Steel	-	-	19	5	1.0	-	-	-	-	-	13	6.3	0.5	-	-	-
1/2 Mo	-	-	29	11	1	-	-	-	-	-	21	6.5	0.5	-	-	-
1 Cr-1/2 Mo	-	-	-	12	1.5	-	-	-	-	-	-	8.0	0.5	-	-	-
1-1/4 Cr-1/2 Mo	-	-	29	11	2	0.6	-	-	-	-	20	8.5	1.0	0.2	-	-
2 Cr-1/2 Mo	-	-	29	11	3.3	-	-	-	-	-	-	5.8	1.3	-	-	-
2-1/4 Cr-1 Mo	-	-	-	13	4	-	-	-	-	-	-	7.0	2.5	-	-	-
5 Cr-1/2 Mo	-	-	25	10	2.5	-	-	-	-	-	17	7.5	1.5	-	-	-
7 Cr-1/2 Mo	-	-	-	8	2.4	-	-	-	-	-	-	5.0	1.0	-	-	-
9 Cr-1 Mo	-	-	-	12	2.5	1.0	-	-	-	-	-	9.0	1.0	-	-	-
12 Cr - Type 410	-	-	33	11	2	1.0	-	-	-	-	20	7.0	1.0	-	-	-
Stainless - 304	-	-	33	18	7.5	2.5	-	-	-	-	25	11	4.0	1.5	-	-
Stainless - 316	-	-	-	25	14	5.0	-	-	-	-	-	15	7.0	2.5	-	-
25 Cr-12 Ni Steel	-	-	-	16	8.0	2.5	-	-	-	-	-	10	4.5	1.5	-	-
25 Cr-20 Ni Steel	-	-	-	18	8.5	2.0	-	0.3	-	-	-	11	6.0	2.0	-	-
Incoloy 800	-	-	-	45	15.0	4.6	2.0	0.8	-	-	-	-	11.0	3.3	1.4	-
Copper Base Alloys																
Copper	6.0	-	-	-	-	-	-	-	3.1	-	-	-	-	-	-	-
Tin Bronze	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Aluminum Bronze	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Ampco No. 18	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Red Brass	9.5	-	-	-	-	-	-	-	8.8	1.0	-	-	-	-	-	-
Yellow Brass	5.0	-	-	-	-	-	-	-	2.0	-	-	-	-	-	-	-
Muntz Metal	-	-	-	-	-	-	-	-	2.0	-	-	-	-	-	-	-
Admiralty	19	2	0.2	-	-	-	-	-	13.0	1.0	.05	-	-	-	-	-
Silicon Bronze	15	-	-	-	-	-	-	-	10	-	-	-	-	-	-	-
70-30 Cu-Ni	-	-	-	-	-	-	-	-	-	-	8.5	-	-	-	-	-
Nickel Base Alloys																
Monel 400	-	-	26	3.0	0.4	-	-	-	-	-	15	3.5	-	-	-	-
Monel K-500	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Nickel 200	-	40	6	-	-	-	-	-	-	13	2	-	-	-	-	-
Inconel 600	-	-	40	12.5	8.0	4.0	1.0	0.5	-	-	30	5	2	1	0.5	0.3
Hastelloy B	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Hastelloy C	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Hastelloy F	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Hastelloy X	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Super Alloys																
Nimonic 80	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Inconel X-750	-	-	-	-	62	30	9.0	-	-	-	-	-	54	23	7.0	-
M-252	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Refractaloy 26	-	-	-	-	45	25	-	-	-	-	-	-	-	-	-	-
A-286	-	-	-	80	27	-	-	-	-	-	-	72	-	-	-	-
Discaloy	-	-	-	-	35	-	-	-	-	-	-	-	30	-	-	-
N-155	-	-	-	-	29	12	-	-	-	-	-	-	22	9.0	-	-
S-590	-	-	-	47	29	15	8.0	-	-	-	-	-	22	10	5.0	-
S-816	-	-	-	-	28	16	13.0	-	-	-	-	-	14	11	3.0	-
Haynes Alloy 31	-	-	-	-	-	-	-	-	-	-	-	-	-	15	4.0	-
Haynes Alloy 25	-	-	-	-	-	-	7.0	3.0	-	-	-	-	-	-	5.0	2.0
Haynes Alloy 21	-	-	-	34	23	14	6.0	-	-	-	-	23	8	-	-	-
16 Cr-25 Ni-6 Mo	-	-	-	35	22	10	-	-	-	-	-	24	13	6.5	-	-
19-9 DL	-	-	76	40	19	8	-	-	-	-	-	32	17	8.0	4.0	-
17-4 PH	-	135	60	-	-	-	-	-	-	125	45	-	-	-	-	-
Other Metals																
Aluminum	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Lead	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Titanium	-	-	-	-	-	-	-	-	38	32	10	-	-	-	-	-
Zirconium	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Tantalum	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-

Ref.: Iron Base Alloys - ASM Handbook, ASTM Bulletins 100, 124, 157, 180, 228; Bulletins Timken Roller Bearing Co. National Tube Co., and Babcock and Wilcox Tube Co.
Copper Base Alloys - ASM Handbook, Bulletins Ampco Metals, Inc., Bridgeport Brass Co., Scovill Mfg. Co., and ASTM Bulletin 181.
Nickel Base Alloys - ASM Handbook, Bulletins International Nickel Co. and Haynes Stellite Co.
Super Alloys - ASM Handbook, ASTM Bulletin 160, Engineering Alloys Digest, International Nickel Co., Haynes Stellite Co., Alleghany Ludlum Steel Co., Cyclops Steel Co., and Crucible Steel Co.

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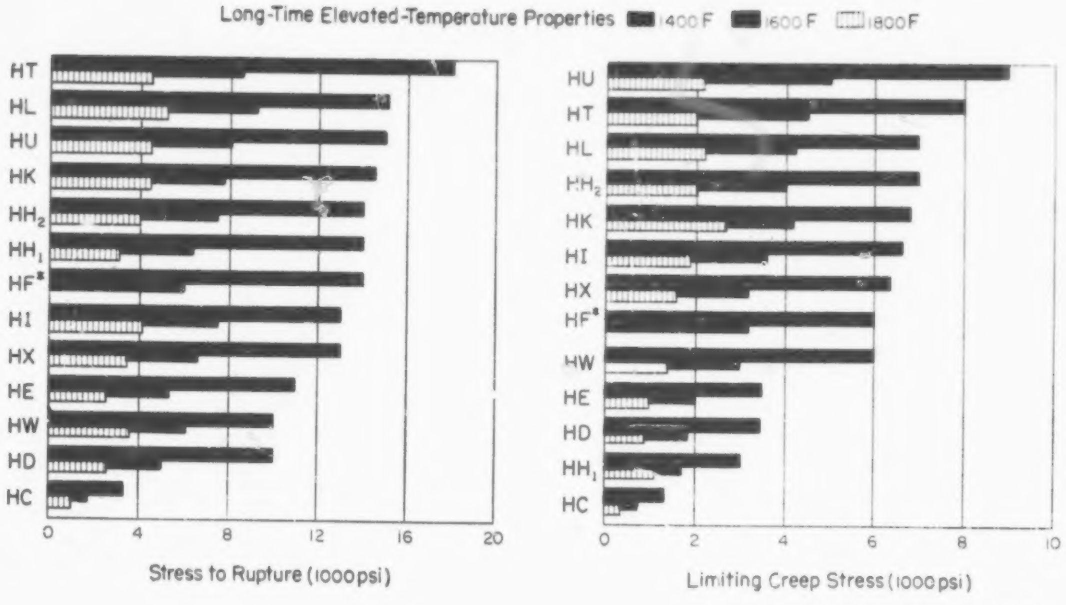

*Not recommended for service above 1800 F.

Figure 7-2. STRESS RUPTURE AND LIMITING CREEP STRESS OF CAST STAINLESS STEELS AT ELEVATED TEMPERATURES
Source: Schoefer, E.A., Machine Design, April 2, 1959
Letters at Left Margins of Graphs Refer to Alloy Designations;
See ASTM Std. A 297

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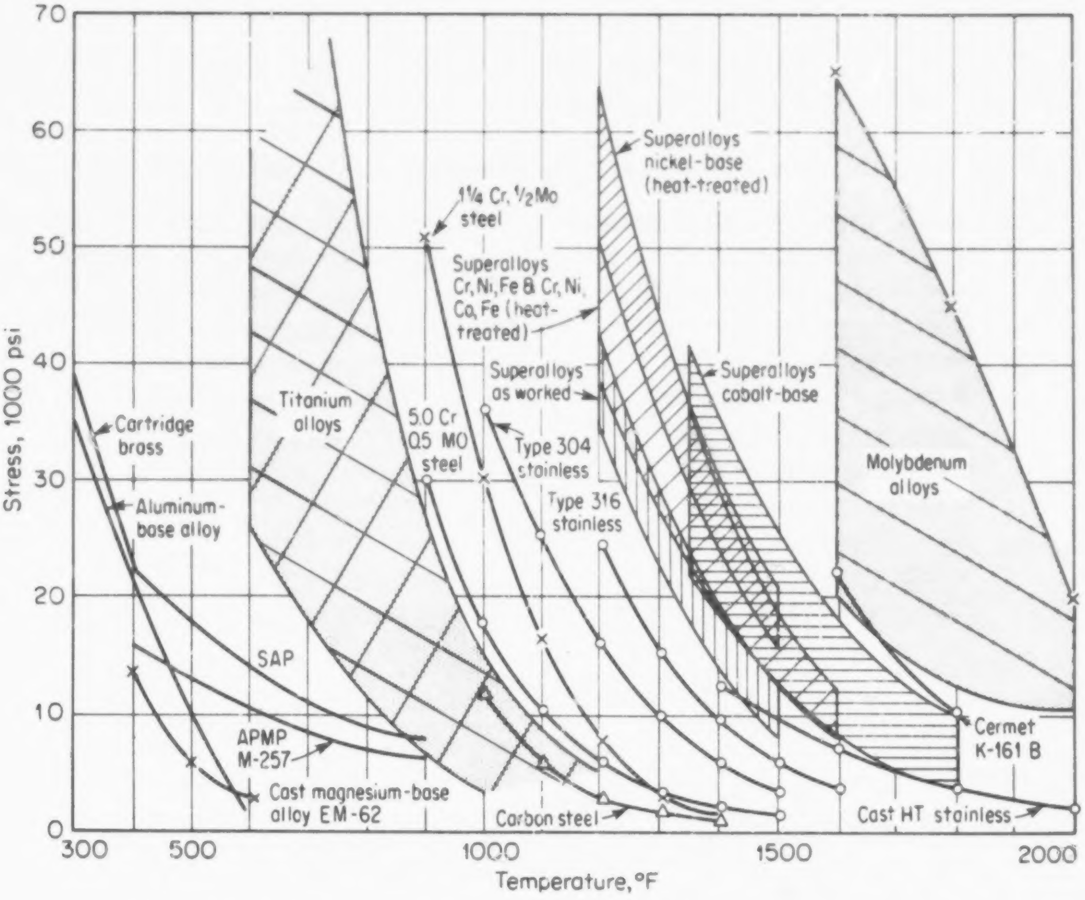

Figure 7-3. STRESS TO RUPTURE IN 1000 HR vs TEMPERATURE

H.C. Cross and W.F. Simmons, ASM Preprint, 1954. See also "Super Alloys", Universal-Cyclops Steel Corp., Bridgeville, Pa.

is commonly used for applications below 1900°F.

7.3.2 Oxidation Resistance of Commonly Used Alloys

The relative oxidation resistance of a number of steels, stainless steels and nickel base alloys is indicated by the data in Table 7-5 and Figure 7-4. It will be seen that high nickel and chromium contents are associated with increased oxidation resistance.

Low alloy steel boiler tubes can be used frequently in heat exchangers because the metal temperature is much lower than the furnace temperature. However, if there is a fouling deposit on the cold side of the tubes, it reduces heat transfer and increases the metal temperature. If this occurs, the rate of oxidation increases and tensile properties are reduced, both contributing to possible tube failure.

7.3.3 Problems of Special Chemical Environments

7.3.3.1 Sulfur Dioxide and Hydrogen Sulfide

The relative effect of SO_2 in air or in air plus water vapor on corrosion of steel at 1650°F is seen from the data in Table 7-6. Sulfur dioxide significantly accelerates corrosion under these conditions.

The relative resistance of various metals and alloys in sulfur containing atmospheres is indicated by the tabulation in Table 7-7.

We have no information on corrosion of metals by SO_3 in flue gas under comparable conditions.

7.3.3.2 Compounds Containing Vanadium, Sodium, Potassium, Sulfur, Molybdenum, or Lead

Compounds containing the above materials are capable of accelerating corrosion of steels, stainless steels and other metallic materials of construction at elevated temperature. Some of these constituents are found in residual fuels. While the concentration in the fuel may be quite low, deposits on metal surfaces in boilers may contain these components in high concentration. Vanadium and sodium react to form products melting as low as 980°F. This liquid and others from the elements mentioned above promote severe corrosion. The mechanism of corrosion appears to be transfer of oxygen to the metal via the concentrated liquid to form a porous metal oxide as the corrosion product. Thus, oxygen is essential to the reaction, and the liquid on the metal surface is not used up in the corrosion reaction. Corrosion by vanadium in fuel oil is illustrated by the corrosion data in Figure 7-5.

Calcium oxide, strontium oxide, and other compounds combine with vanadium oxides and have been added to fuel oils to minimize corrosion.

Table 7-5. MAXIMUM TEMPERATURE WITHOUT EXCESSIVE SCALING IN AIR[a)]

Type of Steel	Maximum Temperature Without Excessive Scaling, °F
Carbon steel	1050
½ Mo steel	1050
1 Cr ½ Mo steel	1100
2¼ Cr 1 Mo steel	1175
5 Cr ½ Mo steel	1200
12 Cr AISI 410	1250
18 Cr 8 Ni AISI 304	1600
18 Cr 8 Ni Nb AISI 347	1600
18 Cr 8 Ni Mo AISI 316	1600
25 Cr 20 Ni AISI 310	2100

a) Data from "Corrosion Engineer's Reference Book", 2nd edition, BIPM, 1965, p. 133.

Table 7-6. EFFECT OF INDUSTRIAL GASES OF DIFFERENT COMPOSITIONS ON GASEOUS CORROSION OF CARBON STEEL AT 1650°F[a)]

Atmosphere	Relative Corrosion
Pure air	100
Air + 2% SO_2	118
Air + 5% H_2O	135
Air + 5% SO_2 + 5% H_2O	276

a) Source: Tomashov, N. D., "Theory of Corrosion and Protection of Metals", The MacMillan Company, New York, 1966, p. 114.

Figures 7-4 and 7-5 follow

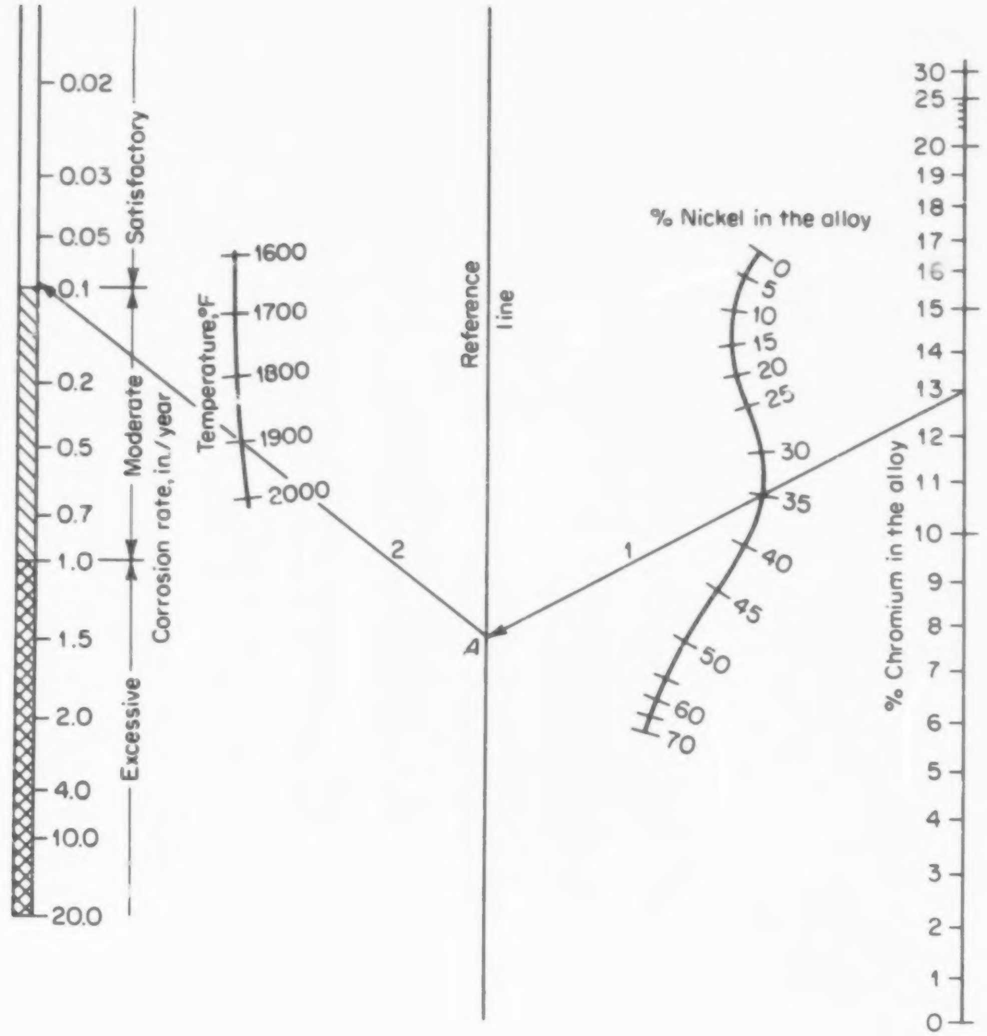

Figure 7-4. AIR CORROSION OF Fe-Ni-Cr ALLOYS

This nomographic chart is based on the 100-hr test data of Brasunas, Gow, and Harder, (Proc. Am. Soc. Testing Mater., 46: 870, 1946) and presents air-corrosion data in concise and convenient form which would otherwise require pages of tables or numerous curves. Loss in accuracy is to be expected in condensing data and, should more precise data be required, the original article may be consulted.

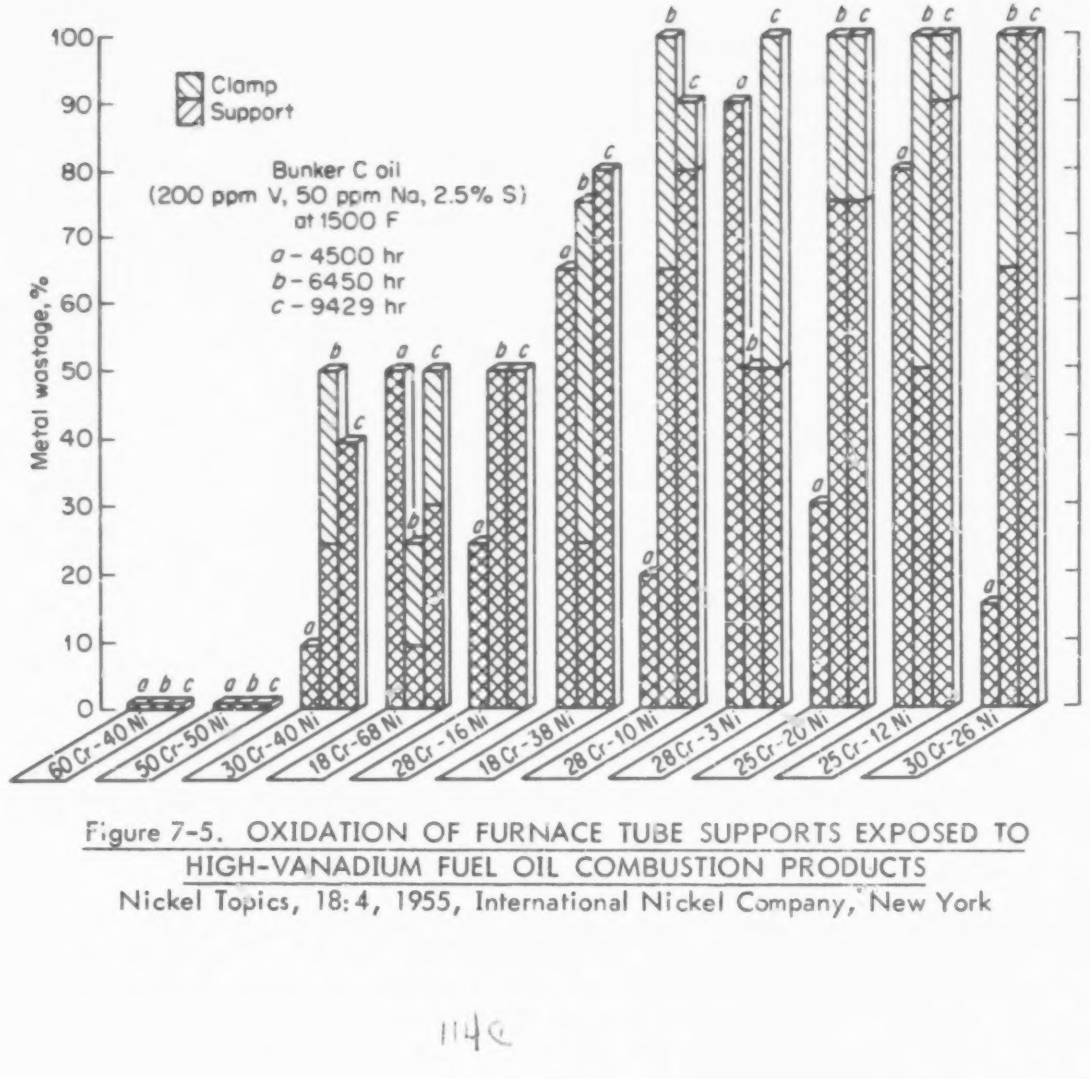

Figure 7-5. OXIDATION OF FURNACE TUBE SUPPORTS EXPOSED TO HIGH-VANADIUM FUEL OIL COMBUSTION PRODUCTS

Nickel Topics, 18:4, 1955, International Nickel Company, New York

Table 7-7. RESISTANCE OF VARIOUS METALS AND ALLOYS TO SULFUR-CONTAINING GASES[a]

Materials listed in order of increasing corrosion resistance

H_2S	S	SO_2
Ni, Co	Ni, Cu	Ni
Mild steel	Mild steel	Fe
Fe	Fe	Cu-10 Mg
Fe-Mn	Fe-14 Cr	Fe-15 Cr
Inconel	Cu-Mn	Ta
Cu	80 Ni-13 Cr-6.5 Fe	Cu, brass
Fe-15 Cr	Mn	Al alloys
Fe-25 Cr	Cr	Mo, W
Cr	Fe-17 Cr	Fe-30 Cr
Fe-18 Cr-8 Ni	Fe-18 Cr-8 Ni	Fe-18 Cr-8 Ni
Fe-22 Cr-10 Al	Hastelloy	Inconel
Cu-10 Mg	Al, Mg	Cu-12 Al
Fe-12 Al, Ni-15 Al		Zr
Ta, Mo, W		
Al, Mg		

a) Source: O. Kubaschewski and B. E. Hopkins, "Oxidation of Metals and Alloys", p. 277, Butterworth and Co. (Publishers), Ltd., London, 1962.

7.3.3.3 Carbon Dioxide and Carbon Monoxide

The presence of carbon dioxide and carbon monoxide in flue gas probably does not accelerate corrosion in most cases with iron base alloys. However, in high CO_2 concentrations or at sufficiently high temperature a CO-CO_2 atmosphere is capable of decarburizing or carburizing steels (and other alloys) by the forward and reverse directions of the reaction

$$C_{(Fe)} + CO_2 \leftrightharpoons 2\,CO.$$

In addition, a CO-CO_2 atmosphere is capable of directly oxidizing iron or reducing iron oxide by reactions such as

$$Fe + CO_2 \leftrightharpoons FeO + CO.$$

An equilibrium diagram for Fe-CO-CO_2 is given in Figure 7-6 which shows the regions of stability of the various species.

7.3.3.4 Chlorine and Hydrogen Chloride

Carbon steel and low alloy steels are not suitable for exposure at furnace temperatures in gaseous Cl_2 or HCl. In Tables 7-8 and 7-9 are listed the suggested maximum temperatures for exposure of various alloys in dry chlorine vapor and dry HCl vapor at elevated temperature. It will be seen that the temperature listed for steel is 400°F in Cl_2 and 500°F in HCl.

Corrosion tests (Figure 7-7) at 900°F with a few percent Cl and steam present indicate severe corrosion of carbon steel and Type 405 stainless steel. Hastelloy C was the most resistant of the alloys tested.

7.3.3.5 Bromine and Hydrogen Bromide

We have no specific information on corrosion by Br_2 or HBr at elevated temperature. Temperature limits may be similar to those for Cl_2 and HCl.

7.3.3.6 Phosphorous Compounds

When phosphorous compounds are burned in the presence of water vapor, the combustion product will condense as a liquid except at very high temperatures. The boiling point of the azeotropic mixture of phosphorous pentoxide and water (92% phosphorous pentoxide) is 1600°F.

Corrosion of carbon steel and Type 316 stainless steel is greater than 3 inches/year in concentrated phosphoric acid (97 to 127%) at 450°F. It appears that a liquid phase of P_2O_5 or other phosphorous compound would probably be very corrosive to most metallic materials of construction at elevated temperature.

Figures 7-6 and 7-7 follow

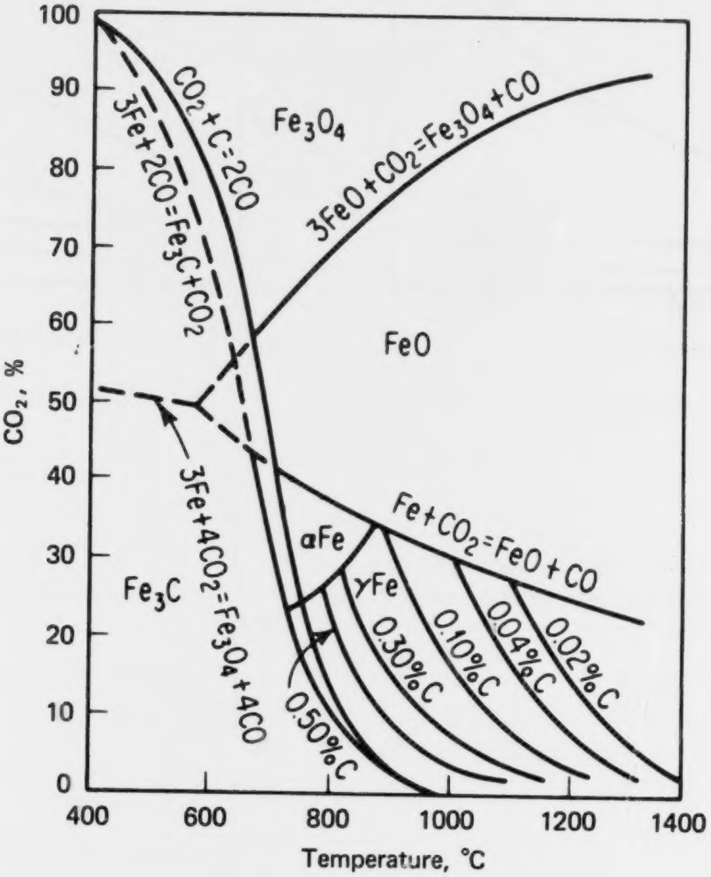

Figure 7-6. EQUILIBRIUM DIAGRAM OF THE Fe-CO-CO_2 SYSTEM

W.D. Jones, "Fundamental Principles of Powder Metallurgy", p. 537, Edward Arnold (Publishers) Ltd., London, 1960

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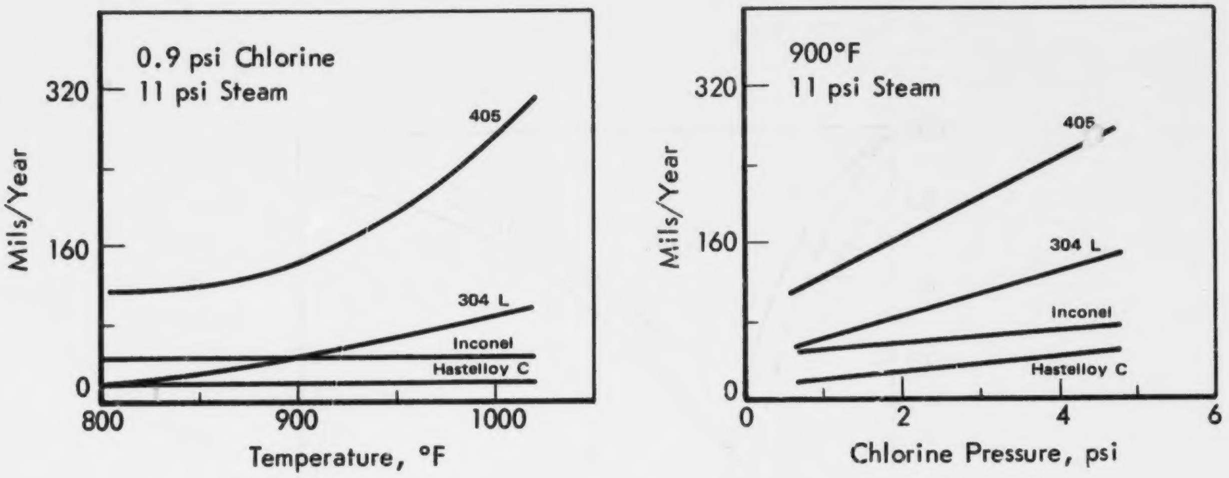


A. Temperature

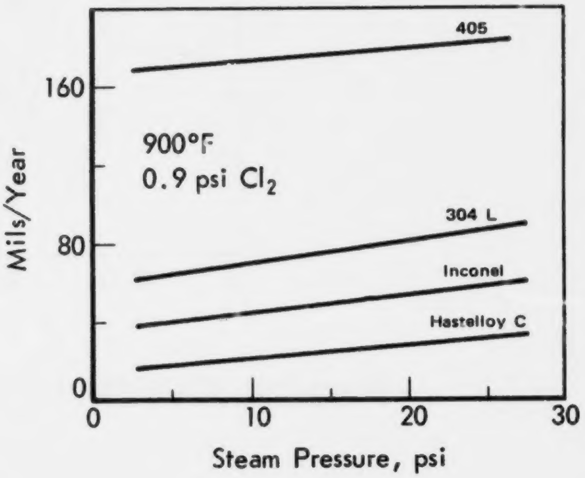
B. Chlorine Pressure

Corrosion rates were calculated assuming the exposure time was the time for chlorination period of single cycle laboratory tests.

Complete cycle consisted of:

1. Oxidation (1.5 hrs, 4 psi O_2, N_2)
2. Chlorination (1.5 hrs, 4 psi O_2, Cl_2, H_2O, N_2)
3. Stripping (2 hrs, 4 psi O_2, 4 psi H_2O, N_2)

Temperature 900°F Except as Indicated
Total Pressure 200 psig
UOP R-8 Catalyst Present



C. Steam Pressure

Figure 7-7. CORROSION BY CHLORINE IN LABORATORY TESTS RELATING TO REFORMER CATALYST REGENERATION

Corrosion is primarily, but not totally, from chlorination portion of cycle. The effect of variables in chlorination portion of cycle on corrosion is illustrated by data above.

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Table 7-8. HIGH TEMPERATURE CORROSION BY DRY HYDROGEN CHLORIDE GAS

Based on data from short term laboratory tests by Brown, DeLong, and Auld (Ind. Eng. Chem., 39, p. 839-44, July 1947).

Material	Suggested Upper Temperature Limit, °F, for Continuous Service
Platinum	2200
Gold	1600
Nickel	950
Inconel	900
Hastelloy B	850
Hastelloy C	850
Type 316 SS	800
Type 304 SS	800
Carbon steel	500
Monel	450
Copper	200
Aluminum	Not suitable at ambient temp[a)]

a) Uhlig, H. H., "Corrosion Handbook", John Wiley and Sons, page 767.

Table 7-9. HIGH TEMPERATURE CORROSION BY DRY CHLORINE GAS

Based on data from short term laboratory tests by Brown, DeLong, and Auld (Ind. Eng. Chem., 39, p. 839-44, July 1947).

Material	Suggested Upper Temperature Limit, °F, for Continuous Service
Nickel	1000
Inconel	1000
Hastelloy B	1000
Hastelloy C	950
Monel	800
Type 316 SS	650
Type 304 SS	600
Platinum	500
Carbon steel	400a)
Copper	400
Aluminum	250b)
Tantalum	300c)
Gold	<250

a) Ignites at 450 to 500°F.
b) Ignites at 270 to 450°F; literature references vary widely.
c) Metals Handbook, ASM, vol. 1, p. 1222.

Figure 7-8 follows

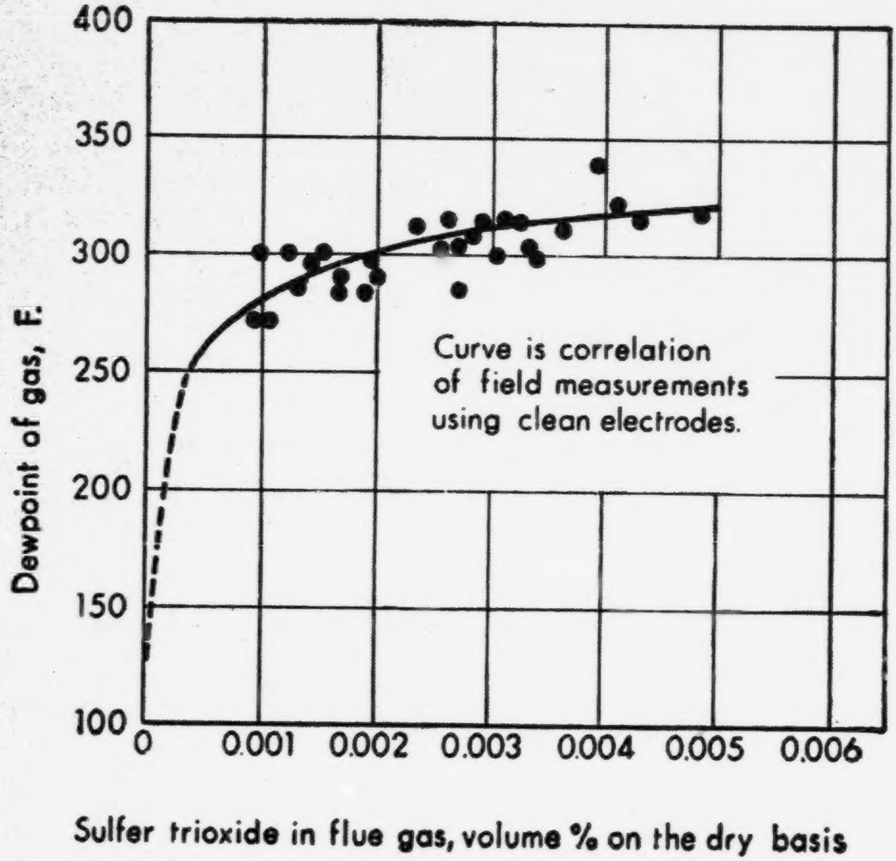

Figure 7-8. WATER DEW POINT WITH SO_3 IN FLUE GASES
Chemical Engineering, Vol. 75, No. 16, p. 168, July 29, 1968

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7.3.4 Further Investigation of Metallic Materials of Construction

The information presented above on metals in furnace environments has served to illustrate the many problems that can be encountered. For any further detailed consideration of this subject it would be desirable to survey published literature, manufacturing practices, and other sources of information on the current state of the art. However, it is likely that laboratory investigation would be required to supplement information available from others. For afterburners, corrosion of metals in atmospheres containing low concentrations of acid gases at furnace temperatures appears to be such an area for possible investigation.

7.3.5 Problems of Heat Recovery Equipment and Cooled Metal Surfaces

Exchangers or steam generators following the afterburner for heat recovery usually require installation of banks of metal tubes that would be exposed to the furnace atmosphere. The temperature of the surface of the tubes would determine the type and extent of corrosion that could occur. High temperature corrosion by various components of the furnace atmosphere has been discussed above.

Another type of corrosion can occur from condensation of a corrosive liquid phase on metal surfaces. The temperature at which such a liquid phase could condense would be determined by the specific components present in the furnace atmosphere.

In any case, prior to shutdown of the furnace it would be necessary to purge all acid gases from the furnace atmosphere in order to prevent corrosion by condensed liquid acidic components on metal surfaces when the furnace is not in operation. Comments on specific acidic components in furnace atmospheres are presented below.

7.3.5.1 Phosphorous Compounds

As indicated above, it appears that concentrated phosphoric acid could condense at temperatures as high as 1600°F.

7.3.5.2 Sulfur Trioxide

Minute amounts of SO_3 normally present in flue gas are strongly absorbed in water droplets so that the corresponding acid condensates are quite concentrated. The sulfuric acid concentration of initial condensate droplets typically falls in the 20 to 80% H_2SO_4 range. Such acid compositions are severely corrosive to steels, stainless steels and many nickel base alloys. The dewpoint of water with very small amounts of SO_3 in flue gas is illustrated by the data in Figure 7-8. These data indicate the dewpoint may be as much as 340°F with SO_3 present. However, other published information indicates condensation temperatures as high as 465°F.

7.3.5.3 Other Acid Gases

The dewpoint of HCl, NO_2, etc., is lower than the acid forming components mentioned above. However, it is necessary to maintain metal surfaces well above such dewpoints to prevent condensation corrosion. For example, in HCl containing a small amount of water vapor it is recommended that the temperature be at least 50°F above the dewpoint to prevent corrosion. This is due to the hygroscopic nature of metal chloride corrosion products.

7.3.5.4 Steel Shell of Afterburner

The afterburner often consists of a steel shell lined with firebrick or castable refractory. As the lining is not totally impervious, furnace gases may be in contact with the steel shell. The temperature of the steel shell is a significant factor in the amount of corrosion that can occur. One way of preventing corrosion by most acidic components is to maintain the temperature of the shell above the condensation temperature of acid gases in the furnace atmosphere (normally above 450°F).

This approach cannot prevent condensation by phosphorous compounds. At one installation, an incinerator was constructed recently that periodically burns phosphourous compounds. Based on a survey of similar furnaces, it appeared that phosphorous compounds would form slags with the refractory. This would tend to seal the surface, preventing exposure of the underlying steel. In addition, acidic phosphorous compounds that penetrated the brick would have been essentially neutralized by reaction with components of the refractory. Thus, it appeared that a steel shell could be used. Similar reactions would not be expected to occur with most other acidic components in the furnace atmosphere.

Another method of preventing corrosion is to have the metal surface at a low enough temperature so that an organic membrane can be used to cover the steel inner surface and thus protect it from the acid gases. For most membranes, this means a temperature of less than 200°F.

Another approach that has been used is to construct a pressure tight steel box from tubes (for heat recovery) and tube fins inside the furnace to prevent flue gas from contacting the exterior.

7.3.6 Thermal Distortion and Thermal Fatigue

Metals, especially the high-nickel stainless steel alloys, have high thermal coefficients of expansion compared with ceramic materials. On the other hand, ceramic parts are often exposed to high temperatures, but supported and restrained by the metal structure. It is of the utmost importance that careful allowance be made for the relative movement between cold and hot parts in such a way that the structure is not damaged. The problem is especially severe during startup conditions, when some parts of the system directly exposed to the flame will heat up quickly, while other parts will heat only after an extended period of operation. Figures 7-9 and 7-1 give data on the thermal expansion coefficients of metallic and ceramic materials at room and elevated temperatures.

Many instances of failures have been encountered where proper allowances for this differential movement have not been provided for. It should be kept in

Figure 7-9 follows

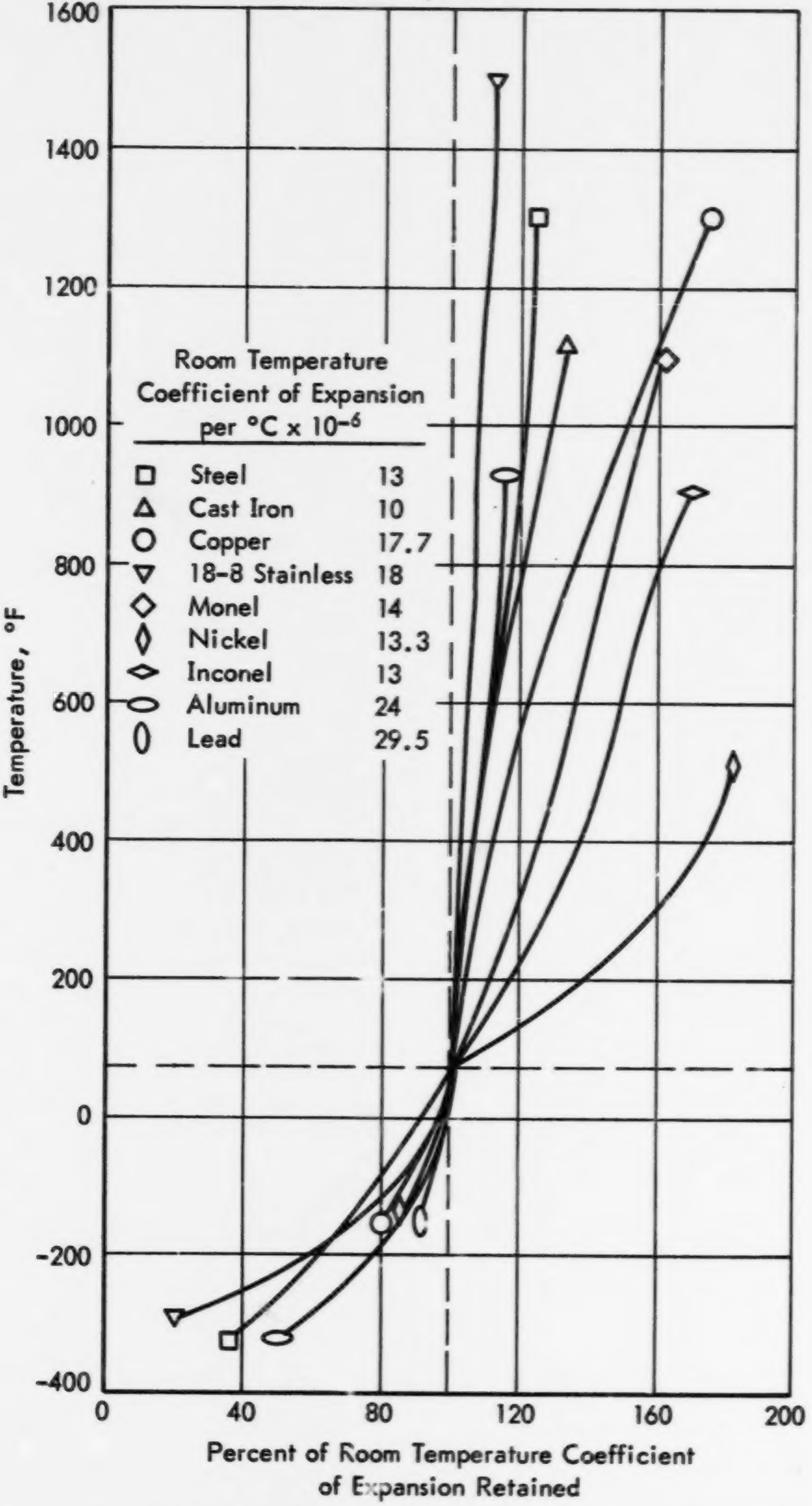

Figure 7-9. THERMAL EXPANSION DATA ON METALS

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mind that restraint on the hot ceramic with a cold structure will often load the ceramic portion of the system in compression; most ceramics are very strong in compression, and will force a containing structure to yield, and the structure will be loose upon cooling. Or if the loading places the ceramic in tension, the low strength and brittle behaviour of the ceramic will lead to the formation of cracks, and possibly to its collapse.

Still further problems arise in the design of metal parts exposed to large variations in temperature. If the thermal expansion is restrained, the structure will warp, twist, and deform. If subjected to repeated cycles of temperature loading in this fashion, the structure will fail from "thermal fatigue." Many instances of warped and cracked metal members in the near vicinity of the flame, even though they may be made of heat-resisting alloys designed for the correct temperature range, attest the severity of the problem.

Chapter 8 AUXILIARIES

The afterburner system is centered around the afterburner itself and its heat exchangers, but consideration must also be given to bringing fumes to the afterburner and discharging the clean flue gas. Also, refractory-lined fume incinerators are heavy and special supporting structures or concrete pads will be required as part of the installation.

Fume collection and delivery to the afterburner is a very important part of the total system and much attention should be paid to its design since both capital and operating costs are sensitive to the manner in which it is handled. Long duct runs can cost more than the afterburner itself and condensation of combustibles can occur even in insulated ducts, causing a fire hazard. Proper hooding is crucial in order to insure that all fumes are collected and that air dilution is kept to a minimum. Once the fume is below 25% of the LEL, all extra air represents wasted fuel in the afterburner. Existing ventilation systems will probably require modification to reduce unnecessary dilution and to vent clean air directly to the atmosphere rather than through the afterburner.

This chapter is not intended to guide the design of the fume collection and delivery system. The reader is directed to "Industrial Ventilation"[8-1] and Chapter 3 of the "Air Pollution Engineering Manual"[8-2] for comprehensive treatments of hood and duct design and blower selection. Forced versus induced draft blowers are discussed in this chapter since afterburner installation will nearly always involve a separate blower. Cost and power requirements are estimated. Cost estimating information is also supplied for ducting, and curves are given for estimating the afterburner weight which must be supported. A brief treatment is given of particle de-entrainment devices installed before the afterburner to reduce the pollutant load entering the afterburner. Some of these same devices could be used on a cooled flue gas stream to remove particulates or acidic components formed within the afterburner (Section 3.1.4.7).

8.1 Blowers

Afterburners are usually supplied with their own blowers which may be either forced draft (waste gas pushed through the afterburner) or induced draft (waste gas pulled through). With induced draft systems, the high temperature of the afterburner discharge usually makes it necessary to blend in cool air immediately before the blower to reduce the effluent temperature to around 600°F. High efficiency heat exchangers (>50% recovery) will achieve this temperature reduction without air dilution. Induced draft fans could be designed to operate at the full afterburner exhaust temperature of 1400°F but this would require special high temperature alloys which would greatly increase fan costs. Also, bearing failures are a problem at high temperature, and even at 600°F, external bearings are utilized.

For a given pressure increase through the blower, the power consumption is proportional to the actual volume of gases handled. The volume at 600°F is twice that at room temperature, and dilution air also increases the volume. Both

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these factors will increase power requirements. Therefore induced draft blowers must be larger (increased capital) and involve higher operating costs than forced draft blowers.

Induced draft systems do offer some advantages, however. The fan sees only a clean, treated stream and there will be no tendency for buildup in the fan blades. Such buildup will occur with forced draft fans in systems where tarry materials are likely to condense or impact on the blades, e.g., smoke house or asphalt blowing offgas. This buildup can cause fires due to friction generated heat or more commonly it may unbalance the fan. Also, leakage of hot and possibly toxic gases can occur in a forced draft system but all leaks will be inward with an induced draft system. This is important since hot gas leakage through a small hole can quickly enlarge the hole and require major repairs.

Normal system pressure drop ranges from 2 in. to 10 in. water column depending on whether the unit is thermal or catalytic, and whether it is fitted with a heat exchanger, and the number of passes in the heat exchanger. The blowers may be generally classified as industrial fans. They are usually fitted with heavy duty radial blades which are simple, robust, and least affected by particulate matter. A number of manufacturers supply them in standard sizes to 50,000 cfm capacity with static pressures from 1 in. to 20 in. water column. Normal maximum operating temperature is 350°F, but fans may be modified for operation up to 900°F by fitting cooling wheels and loose fitted water cooled bearings. Blowers are available with higher pressure ratings but are in a different classification.

Power requirements for blowers are proportional to the actual volume moved and to the total static pressure increase. Plots of brake horsepower against cubic feet per minute at pressures from 2 in. to 28 in. water column are shown in Figure 8-1. An efficiency of 60% has been assumed. When sizing the motor for a blower, care must be taken to insure that the worst conditions have been allowed for. A fixed rpm blower connected to a system of ducts at a uniform temperature throughout, will deliver an actual volume which will be approximately constant regardless of the temperature level. The pressure developed by the blower and the pressure drop through the system are both proportional to gas density and both will decrease as the temperature rises (for constant ACFM). Thus power requirements, which are proportional to the volume handled and the pressure developed will decrease as the temperature rises, and maximum power will be needed on a cold startup. An afterburner installation is somewhat more complicated since the temperature of the waste gas from the process may vary as well as the temperature of the afterburner, and the blower may be either before or after the afterburner. Pressure drop calculations must be made for each part of the equipment (ducting, afterburner, heat exchanger, etc.) for startup conditions to determine maximum motor horsepower needed. Electrical power usage during normal operation is calculated from the operating conditions. In some installations power is conserved during startup by fitting vane control inlets to the blower which reduce the volume of air handled. This same system can be used to reduce power consumption whenever the process is operating at reduced flow rates.

Figures 8-1 and 8-2 follow

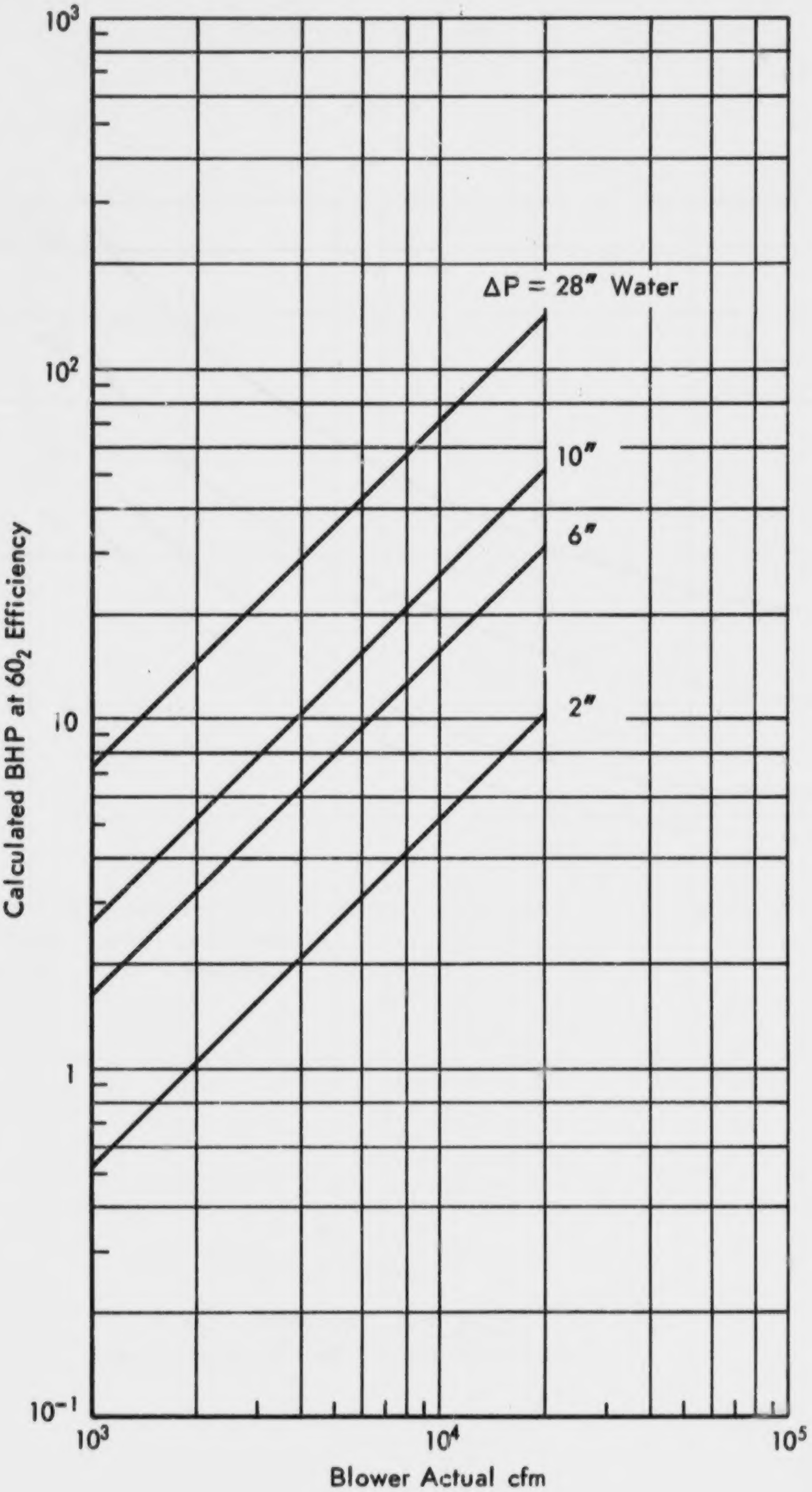

Figure 8-1.

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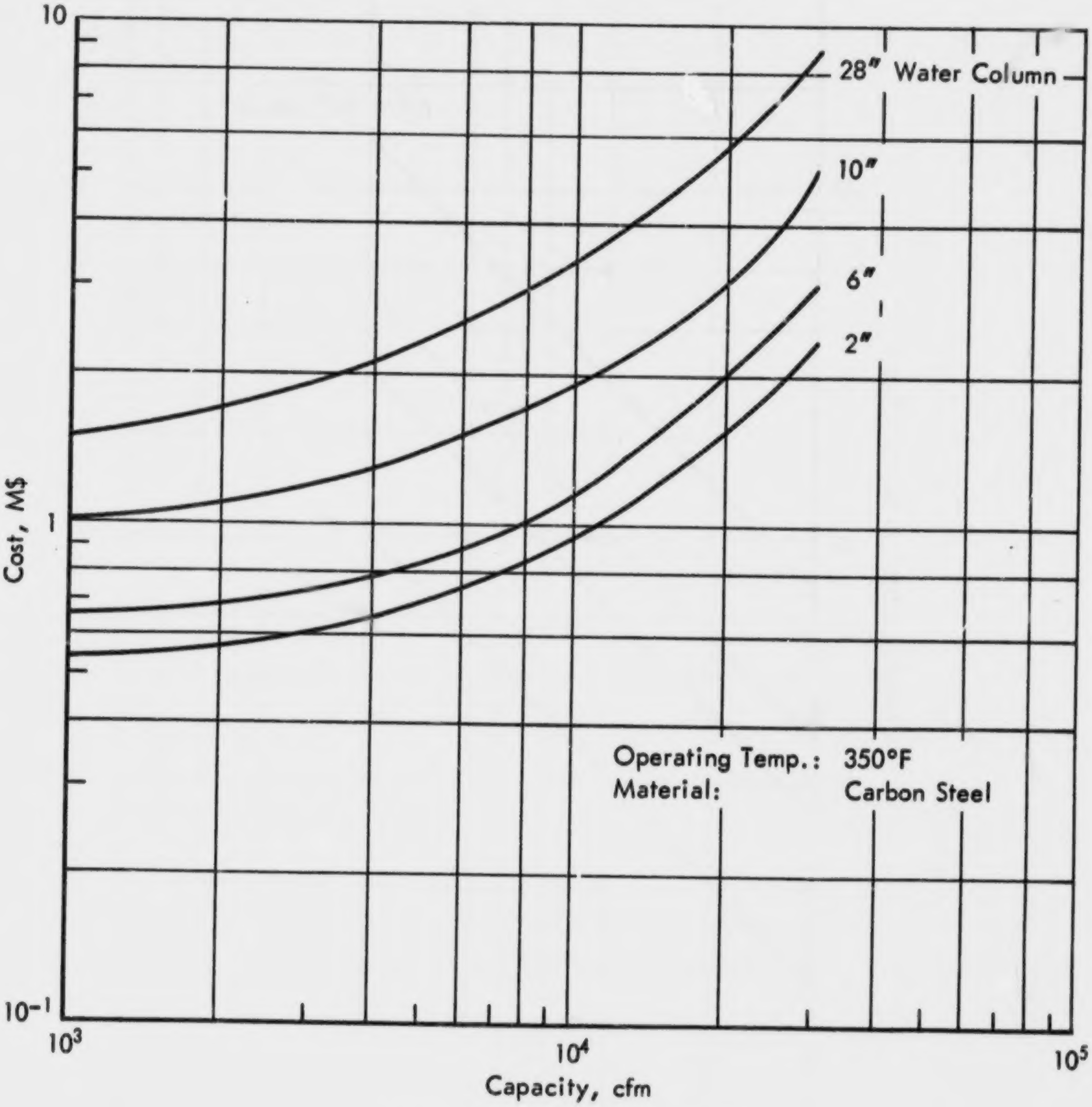

Figure 8-2. BLOWER COSTS
Costs Include Motor, Belt Drive and Guard

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A comprehensive treatment of blowers is given in References 8-1 and 8-2, and much information is available from manufacturers' brochures.

Typical blower costs complete with motor and drive are shown in Figure 8-2, for a system operating at 350°F. In addition to normally encountered pressures of 2 in., 6 in., and 10 in. water column, a plot for 28 in. pressure drop has been included. This indicates the cost penalty incurred when mixing is improved at the expense of pressure drop through the afterburner. Electricity costs will also increase in proportion to the pressure drop. For example, at 10,000 CFM and 2 in. pressure, the annual electrical cost would be $240 per year, whereas if the pressure were increased to 28 in., the cost would be $3300 (based on 6000 hrs/hr and electricity at 1c/KWH).

8.2 Ducts

Ductwork requirements depend largely on the type of installation, the number of units served, and on the location of the afterburner in relation to the process equipment. The ducts are often insulated to prevent condensation of volatile material on the inside. This type of condensation can be a serious fire hazard, and even where the ducts are insulated, it is a wise precaution to install access panels for inspection and clean-out if it should be required.

Most systems operate below the 25% LEL, but for those situations where this concentration is exceeded, provision must be made to prevent flashback. This can be done by providing high velocity sections in which the waste gas velocity is higher than the flame propagation velocity. If concentrations are above the HEL, fumes may be ducted without the need for costly air dilution. (Fuel must be burned to heat up the dilution air to 1400°F.) However, it is essential that air be excluded at all points between the fume generating process and the afterburner. An alarm system and/or careful maintenance are required for rapid detection of leaks.

The cost of ductwork is highly variable, but the following figures may be used as a guide for galvanized carbon steel in relatively short runs:

Size Diameter, in.	Ga.	Duct, $/ft	Insulation, $/ft
6	24	4.00	2.30
9	22	7.40	3.50
12	22	9.80	4.70
15	22	12.20	5.90
18	20	18.00	7.00
24	18	29.60	9.40

Stainless steel ducts would cost approximately 2.5x galvanized ducts.

Methods of calculating pressure drop in ducts, the design of hoods for providing ventilation for different types of equipment and general information on various hydrocarbons, including flammability limits, are covered in some detail in References 8-1 and 8-2.

8.3 Supporting Structures - Afterburner Weight

Installation costs are extremely dependant on the supporting structure required for the afterburner. Figure 8-3 shows the range of equipment weights obtained from manufacturers' data. If the afterburner can be mounted at ground level on a concrete pad, the weight will have little influence on the installation cost. However, afterburners are frequently mounted on the roof of a building. Roof mounting often allows a minimum of ducting since the afterburner can be located directly above the fume generating process. Also roof mounting saves valuable (or unavailable) space within the building and eliminates the need for a tall stack to discharge flue gases above the building. If the roof itself is not strong enough to take the additional load (very common in old buildings), a special supporting structure with expensive steel work will be needed. On a one-story building, it may be fairly easy to run columns to the ground and the additional expense for supports will not be high. However, on a multistory building, costs may be prohibitive and special lightweight designs or ground level mounting will be required. Thus, it is important to consider the weight factor and the cost of supporting structures at the time of the design.

8.4 De-Entrainment

Carryover of particulate matter, both liquid and solid, may add substantially to the load on an afterburner. Solid particles may not burn or may not have sufficient residence time at temperature in the afterburner to be destroyed (Section 3.1.4.6). Excessive liquid carryover of a combustible material may result in excessive heat release in the afterburner with combustion chamber temperature exceeding the design value. This will result in shutting down the unit and/or damaging the installation. It is therefore necessary to provide equipment for removing these particles before the waste gases reach the afterburner. Typical examples are the removal of chaff from coffee roasting off-gases and removal of entrained hydrocarbon mists from asphalt blowing.

There is a large number of different types of equipment for removing particles and mists and there is extensive literature on the subject. References 8-4 through 8-13 cover a representative selection which give general information of the equipment and design criteria. Chapter 4 of Reference 8-2 also covers this subject.

Figure 8-3 follows

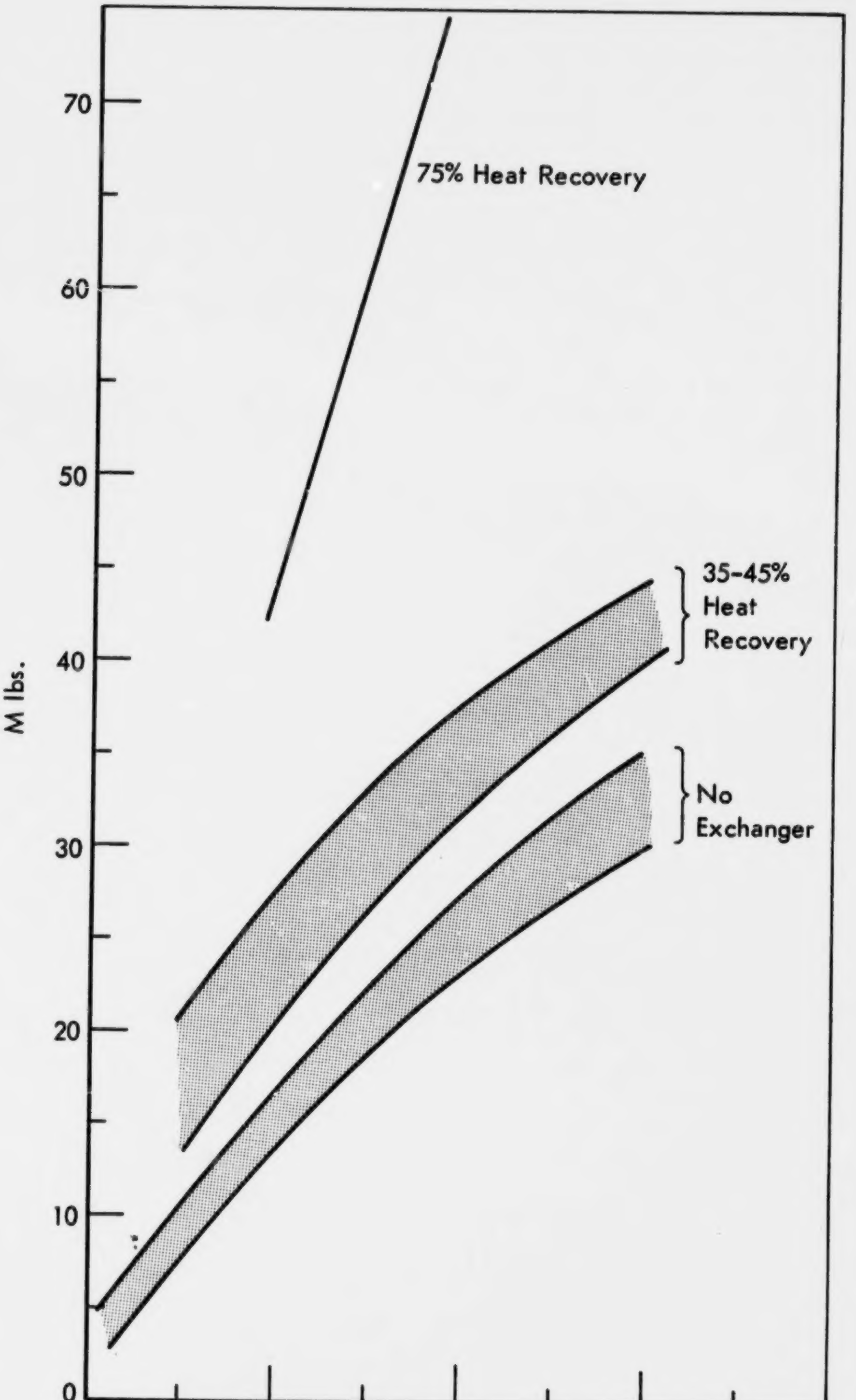

Figure 8-3. AFTERBURNER WEIGHTS - THERMAL

Based on Manufacturers' Information

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Chapter 9. CONTROL AND SAFETY

Failure of control instruments causing major or nuisance shutdowns is one of the most common complaints from the operators of fume incineration equipment. In other cases improperly installed temperature sensing elements or inadequate control systems have resulted in excessive fuel consumption, large temperature variations, and operation at temperatures too low to give acceptable pollutant destruction. Also, proper use of process controls will enable the user to operate drying ovens and similar equipment so as to obtain higher solvent levels in the fume stream, thereby reducing fuel costs. For these and other reasons, it is important that an afterburner system be chosen which includes effective and reliable process controls.

This chapter is intended to guide the selection of control and safety systems for afterburners. Descriptions of the control equipment generally available and typical control systems are given. Recommended practices in regard to control and safeguards are discussed in detail. Included are sections on combustion safeguards, temperature controls, and vapor concentration controls. This chapter is intended to supplement, not supplant, the recommendations of national approval bodies and current Federal, state and local codes and regulations. All process controls for afterburner systems should be approved by local authorities having jurisdiction and the insurance underwriter.

9.1 Combustion Safeguards

9.1.1 National Approval Bodies

There are several bodies in the United States that make recommendations and/or approve equipment and systems relating to safety and fire protection. These bodies include both non-profit organizations and groups sponsored by insurance underwriters. A brief description of the major approval bodies follows.

1. Underwriters Laboratory, Inc.
   207 East Ohio Street
   Chicago, Illinois 60611

Underwriters' Laboratories, Inc. (UL) is a non-profit laboratory for the testing of equipment and systems. They are primarily concerned with the manufacture of equipment in accordance with standards of safety. Once an item or system is approved, they are concerned only that the item continues to be made to the same specifications as the original that was tested.

2. Factory Mutual System
   1151 Boston-Providence Turnpike
   Norwood, Mass. 02062

Factory Mutual System (FM) is the testing laboratory of nine fire insurance companies which specialize in insuring manufacturing plants against fire, explosions, and other losses.

FM publishes many of their findings in manuals and bulletins. They also publish a list of approved devices for industrial fire protection. They consider each system or installation on its own merits before approval is given.

3. Factory Insurance Association
85 Woodland Street
Hartford, Conn. 06102

The Factory Insurance Association (FIA) is composed of member stock insurance companies. FIA is concerned with all phases of fire protection and other perils that are insured by its member companies.

FIA generally, accepts devices that are approved by Underwriters Laboratories, Inc. However, they approve or disapprove each system or installation individually.

4. National Fire Protection Association
60 Battery March Street
Boston, Mass. 02110

The National Fire Protection Association is a non-profit voluntary membership organization with over 200 national trade and professional organizations as Organization Members and more than 24,000 Associate Members in industry, fire departments, government, etc.

All NFPA Official Codes, Standards, and Recommended Practices are published as the National Fire Codes. These standards are used as the basis of local codes and ordinances or building and fire regulations and are widely adopted by industry as a basis of good practice.

It should be noted that for a system to be approved by one of the national agencies, it should be constructed of approved devices where they are available. However, the use of devices which bear an approval label does not necessarily mean that the system is approved. Approval of the system is made on an individual basis.

9.1.2 Recommended Practices

The recommended practices of the national agencies for the installations of combustion safeguards for afterburners are basically the same as those set forth for single burner boilers and furnaces. Although there are some minor differences in definitions and recommendations of the various agencies, on most points there is agreement. All of the approval bodies require a pre-ignition purge to clear any combustible gases from the burner. However, the recommended purging rate varies from 50 to 70 per cent of the maximum air flow and the minimum recommended purge volume ranges from four to eight air changes. A purge timer is required with an interlock that will not allow the fuel valves to open until the purge period has been completed. In the case of afterburners, facilities should allow the pre-ignition purge to be made with fresh air rather than the vent stream to be burned, especially if material may have condensed in the duct during the period the equipment was idle.

An interlock to insure the presence of combustion air is recommended. This is normally a pressure switch plus an electrical interlock to the fan motor starter. Loss of either power to the fan motor or loss of air supply pressure (this could be either positive or negative pressure depending on whether the unit has a forced or induced draft fan) will cause the fuel valves to close. If the afterburner uses the vent stream for the combustion air supply, then the interlocks would be on this stream.

The recommendations for fuel gas valves vary somewhat as shown in Figure 9-1. FM recommends a manual opening safety shutoff valve for manual and semi-automatic burners. NFPA allows automatic opening, spring closing safety shutoff valves but recommends that two safety shutoff valves be installed in series with an automatic closing, spring opening vent valve between them to give a double block and bleed arrangement. They further recommend a position indicator and interlock on the valve nearest the burner. This interlock prevents initiation of the firing sequence until this valve is closed. FIA recommends a double block and bleed arrangement similar to NFPA except that the upstream valve shall be a manual opening safety shutoff valve and the position interlock is not required. For gas to the pilot NFPA and FIA recommend a double block and bleed arrangement with automatic safety shutoff valves. FM allows a single safety shutoff valve.

Any of the above systems should provide safe operation if the equipment is inspected regularly and kept in good condition. However, the FIA and NFPA system provide redundance for added protection. The final choice should, of course, comply with local fire regulations.

For oil fired burners a single automatic opening, spring closing, safety shutoff valve in the oil supply line is recommended by all agencies. Provisions should be made for heating and recirculating heavy oils.

Fuel pressure supervisory switches are recommended. Gas burner installations should have both high and low pressure interlocks with the safety shutoff valves. Oil burners normally require only the low pressure interlock, but other interlocks may be needed on the oil temperature and atomizing steam or air.

All agencies recommend the installation of an approved non-recycling flame safety shutdown system which will stop the fuel flow within 2-4 seconds if the flame is extinguished. The system may detect the presence of a flame by means of a flame rod, or other device sensitive to visible, infra-red or ultra-violet flame energy. The merits of each method will be discussed later. In addition to stopping the fuel on flame failure, the flame safety shutdown system should include timers which will close and lockout the fuel valve if the pilot and/or main flame is not established in a reasonable time.

9.1.3 Firing and Light-Off Options

Many options are covered by recommendations of the FM, FIA, and NFPA. Gas only, oil only, or both gas and oil can be used for fuel to the main burner. The fuel selected will probably be based on the economics but the use of a single fuel, preferably gas, will result in a simpler control system that will be easier to maintain.

The degree of automation available for lighting off the burners ranges from a permissive system where the operations are performed manually to one where all the operations are automatic after the start button is pressed. The definition of these systems as supervised manual, semi-automatic, or automatic varies with the recommending agencies. The user should select the system that best fits his requirements keeping in mind the frequency of startups and the increased maintenance that the more automated systems will require. Any system selected should have a continuous pilot and should not attempt to relight on a flame failure (be non-recycling).

9.1.4 Equipment - Flame Detectors, Shut Off Valves, Pressure Switches

9.1.4.1 Flame Detectors

A flame detector is an essential item in the combustion safeguard system. Most practical flame detectors use one of two basic principles. One type uses a flame rod in actual contact with the flame and employs the rectification of an electric current by the ions in the flame to determine the existence of the flame. The other type depends upon the detection of radiation from the flame. The radiation wave length detected can be in the infra-red, visible or ultra-violet region depending upon the design of the unit.

The flame rod has a fast response. However, its direct contact with the flame results in oil and carbon deposits on the rod which limit its usefulness. They must be cleaned and replaced quite often and, in some types of fuels which have a low resistance flame, the low output of the flame rods results in nuisance shutdowns.

The flame detectors based on radiation do not come in contact with the flame but they must have a clear viewing port so they can see the flame. The detectors are also limited to an ambient temperature of approximately 200°F and may require cooling in some applications. Photocells which sense radiation in the visible region are limited in their ability to reject radiation from the refractory. Infra-red detectors overcome the problem of hot refractory but they can be fooled by seeing shimmering hot air. Ultra violet detectors are reliable in applications to both gas or oil flames. An additional safety feature is available on some ultra-violet detectors that interrupts the view of the flame about 20 times per second. The circuit will re-establish only if ultra-violet radiation still exists. The circuit, in effect, proves to itself 20 times per second that a flame exists.

Figure 9-1 follows

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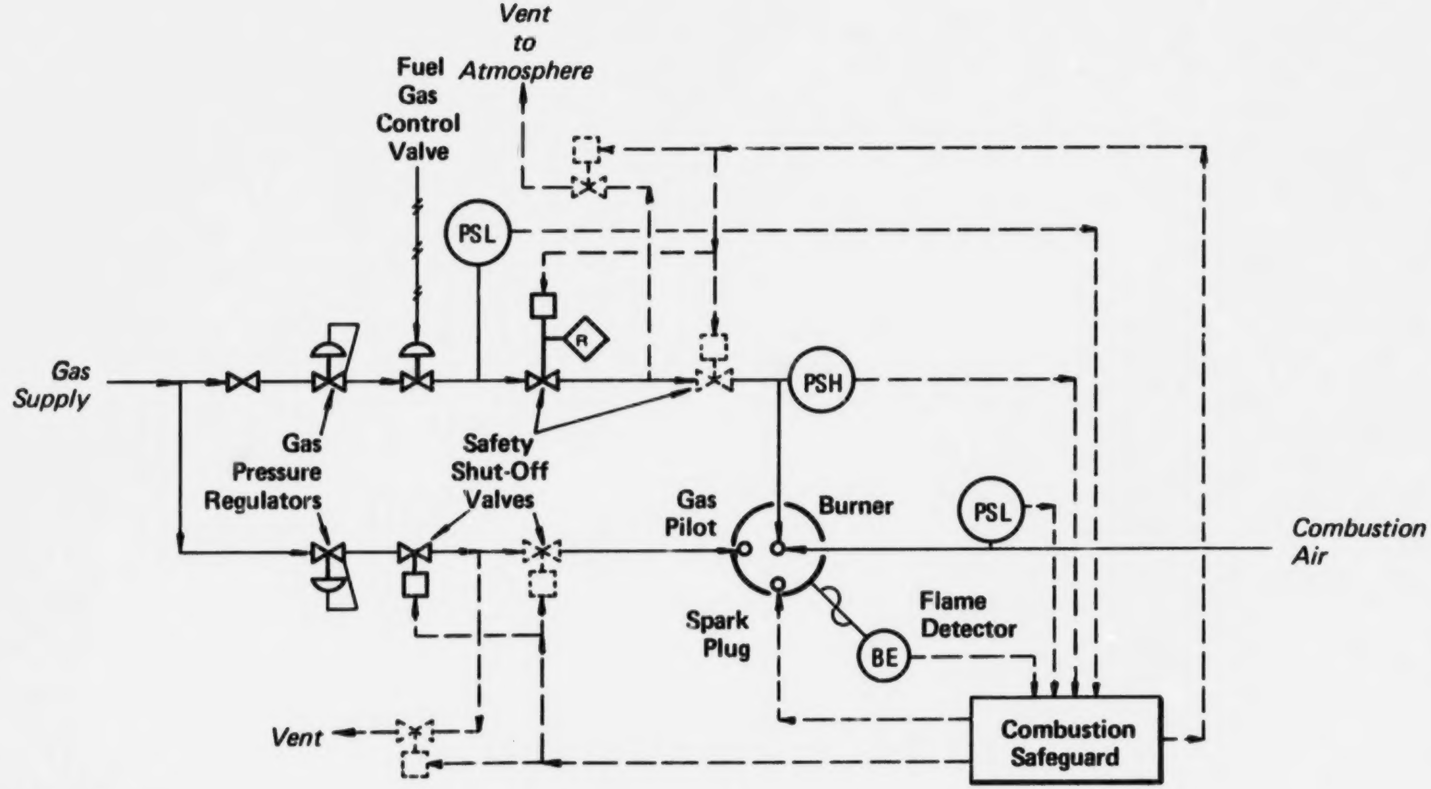
Vent
to
Atmosphere
Fuel
Gas
Control
Valve
PSL
R
Gas
Supply
PSH
Gas
Pressure
Regulators
Safety
Shut-Off
Valves
Gas
Pilot
Burner
PSL
Combustion
Air
Spark
Plug
Flame
Detector
BE
Vent
Combustion
Safeguard

PSL	Press. Switch-Low
PSH	Press. Switch-High

Notes:

1. Valves Shown in Solid Lines are Required by FM.
2. Additional Valves Shown in Dotted Lines Required by FIA and NFPA.
3. NFPA Required Interlock Switches but Not Manual Reset.

Figure 9-1. TYPICAL COMBUSTION SAFEGUARD ARRANGEMENT FOR A SINGLE GAS BURNER

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9.1.4.2 Shut Off Valves

There are several types of shut-off valves used in combustion safeguard systems depending upon the requirements of the approving authority and the degree of automation of the system. The safety shut-off valves required by FIA, FM, and NFPA are electrically operated valves which shut-off the fuel automatically when de-energized. They must be constructed so that they cannot be blocked open and they must be closed by a spring that maintains a minimum force of 5 pounds on the seat. On the valves nearest a gas burner, NFPA requires position switches on the stem for use in the safety interlock system. The manual reset safety shut-off valve is designed so that it can be opened and reset manually only after it has been energized. It closes automatically when the circuit is deenergized regardless of the position of the reset handle. All safety shut-off valves should be approved by the appropriate authority.

9.1.4.3 Pressure Switches

Various switches which sense air flow, temperature, fuel pressure, etc. are necessary to complete the combustion safeguard system. The switches which meet the standards of the UL and/or FM are approved by these bodies.

9.1.5 Typical Gas Fired Single Burner Combustion Safeguard System

The system shown in Figure 9-1 is typical. The recommendation of the appropriate authority should be considered for a specific system.

9.2 Temperature Control

9.2.1 Control Concepts

9.2.1.1 Oxidation Zone

The obvious need for temperature control in the combustion area occurs because a minimum temperature must be maintained in order to properly oxidize the fumes. Maintaining temperatures much in excess of this minimum results in higher fuel costs and also possibly higher maintenance costs. There are also other requirements which must be considered when a temperature control system is selected. In afterburners which operate at high temperatures the refractory may not be able to stand the shock of a rapid change in temperature. In this case a temperature programmed startup system may be required with the temperature control point raised slowly to the final operating point. This can be done either automatically or manually if the temperature span of the controller is broad.

Some burners may have a limited turndown and at low fume loads may result in higher than desired fire box temperature. This problem can usually be handled by split range control of the burner valve and another valve which admits dilution air to the afterburner. As the load on the afterburner is reduced, the controller reduces the fuel to the burner until the minimum allowable setting is reached. The dilution air valve then is opened and additional cool air mixed with the incoming fumes to maintain the firebox temperature at the control point.

9.2.1.2 Shutdowns and/or Alarms

In addition to the normal temperature controls there should be a separate thermocouple and high temperature shutdown that will shut-off the fuel valve and vent the fumes in case the firebox temperature exceeds the safe limit. A low temperature alarm should also be considered if the fumes cannot be properly oxidized below certain temperatures.

Whenever the burner is shutdown either manually or automatically, the fumes should be shut-off and fresh air admitted to the burner. The fan should continue to run long enough to purge the fumes from the combustion chamber, but not long enough to chill the refractory if this can cause damage. A timer could be considered to provide the purge automatically.

9.2.1.3 Inlet Fume Temperature

In some afterburner installations the combustion products pass through a heat exchanger to preheat the incoming vapor stream. This reduces operating costs but unless precautions are taken a reduced flow of fumes to the afterburner can cause overheating, possible preignition, and damage to the heat exchangers and/or ductwork. If condensation has previously taken place in the ductwork the increased temperature at reduced flows may cause re-evaporation and result is an explosive mixture in the ductwork. Control of the temperature of the incoming fumes will usually prevent trouble. In certain installations, it may be necessary to provide a controlled bypass around the exchanger in order to adequately control the preheat temperature (see Chapter 6).

9.2.2 Burner Control

9.2.2.1 Recommended Practices

The control of the temperature of the combustion area of the afterburner will normally be accomplished by modulating the flow of fuel to the burner. It is of the utmost importance that measurement of the temperature to be controlled is properly made. The temperature measuring point must be located where the temperature is representative of the mixture of fumes and the combustion products of the burner. This will require that the probe be in a location where the gases are well mixed and are moving at a velocity high enough to insure good heat transfer. The probe should not be mounted where it is subject to flame radiation or impingement or in a dead spot where the gases are essentially stagnant.

9.2.2.2 Equipment - Thermocouples, Controllers, Control Valves, Overtemperature or Safety Devices

Thermocouples

Most afterburners will operate in a temperature range of 850-1500°F. In this range, base metal thermocouples are the most inexpensive and trouble free means of measuring temperature. Iron-Constantan (Type J) couples are inexpensive and are good to 1400°F. At temperatures up to 2300°F, Chromel-Alumel (Type K) couples should be used. The thermocouples should be installed in wells to protect them from oxidation and/or corrosion.

Controllers

The selection of the temperature controller will depend upon economic factors and the required precision of the control. Regardless of the type selected, the controller, or the transmitter if one is used, should have upscale thermocouple burnout protection. This means that in case of a faulty thermocouple circuit, an abnormally high signal will be seen by the controller which will shut-off the fuel to the burner.

The most simple form of controller is the electric temperature controller with relay output. This controller accepts an input directly from the thermocouple and operates relay contacts which cause an electric motor to open or close the fuel valve. Since the thermocouple connects directly to the controller, the signal level is very low and the lines must be kept short to avoid losses and interference. The form of control is usually simple three position control. If the temperature is too low, a relay closes which opens the fuel valve. Another relay operates and closes the fuel valve if the temperature is too high. The "dead zone" between the two relay operating points is adjustable as well as the proportional band or the speed that the valve operates. This controller provides a reliable system at a minimum cost and requires only electric power for operation. However, because of the "dead zone" necessary to prevent continuous cycling of the valve, the temperature will float between the high and the low setting. This is not a problem unless very precise temperature control is required.

If the afterburner is installed in a plant which has conventional pneumatic or electronic process control instrumentation, it may be desirable to use the same type of equipment for controlling the afterburner temperature. The temperature can be controlled more precisely and spare equipment for maintenance may be reduced. However, more equipment is usually required and the cost will be greater. Oil-free air at 20 psi will be needed to operate the valve. A transducer is normally required to convert the thermocouple signal to a standard pneumatic or current signal that the process controller will accept. Since the pneumatic or current signal is a high level signal, it can be transmitted a considerable distance to a remote controller if desired. The output signal from the controller is a standard pneumatic or current signal which ultimately operates a pneumatic control valve in the fuel line.

Control Valves

The type of control valve will be dictated by the controller selected. The electric temperature controller with the relay output requires a reversible proportioning motor acturator on the control valve. The use of the proportioning motor actuator is limited to valves 3-4 inches and under with low pressure drops. The stroking speed is relatively slow compared to the pneumatic operated valves. However, these are not serious disadvantages in installations of one or two burners.

Both the pneumatic and electronic process controllers require a pneumatic valve operator. An electronic valve positioner or transducer is also required for the valve operated by the electronic controller.

Overtemperature or Safety Devices

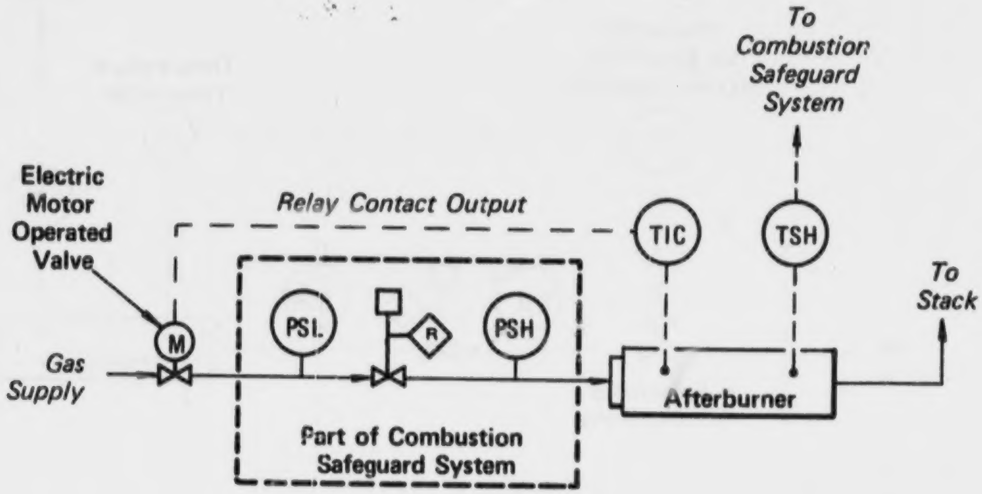
The sensing points of any combustion area temperature alarms should be located just as carefully as the points for the temperature control. Again, thermocouples are probably the best choice for sensors for low cost and reliability at elevated temperatures. Reliable and inexpensive solid state switches are readily available which will accept a thermocouple input. These switches can operate high or low temperature alarms or be connected to the combustion safeguard system to shutdown the burner.

9.2.2.3 Typical Burner Controls and Alarms

The system shown in Figure 9-2 is typical of a system that will provide reliable temperature control at a minimum cost (approximately $750).

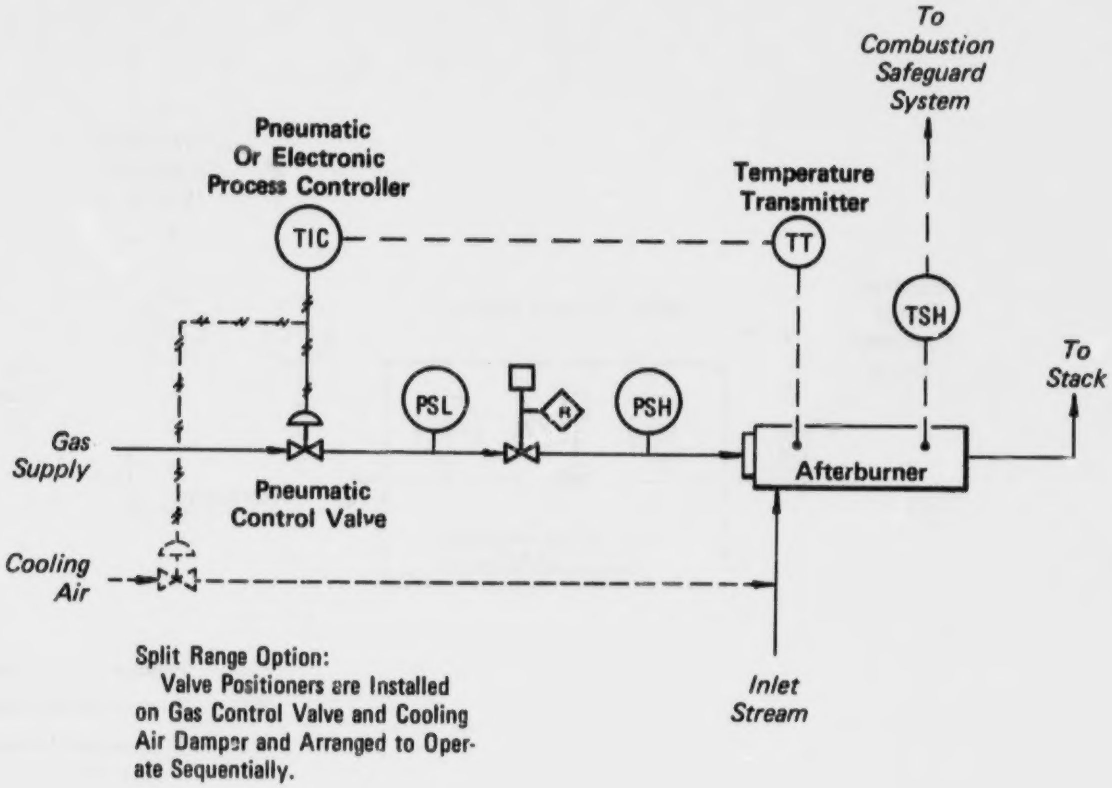
A more precise and complex system is shown in Figure 9-3. This system might be installed as part of a large plant where instrument air is available and remote monitoring may be desirable. The cost for a pneumatic system would be approximately $1200 and for an electronic system about $1400.

Figures 9-2 and 9-3 follow


To Combustion Safeguard System
Electric Motor Operated Valve
Relay Contact Output
TIC
TSH
To Stack
M
PSL
R
PSH
Gas Supply
Afterburner
Part of Combustion Safeguard System

PSL	Pressure Switch-Low
PSH	Pressure Switch-High
TIC	Temperature Indicating Controller
TSH	Temperature Switch-High

Figure 9-2. MINIMUM COST TEMPERATURE CONTROL SYSTEM


To Combustion Safeguard System
Pneumatic Or Electronic Process Controller
Temperature Transmitter
TIC
TT
TSH
To Stack
PSL
H
PSH
Gas Supply
Afterburner
Pneumatic Control Valve
Cooling Air
Inlet Stream

Split Range Option:
Valve Positioners are Installed on Gas Control Valve and Cooling Air Damper and Arranged to Operate Sequentially.

Figure 9-3. TEMPERATURE CONTROL SYSTEM USING PROCESS CONTROL INSTRUMENTATION

Alarm points or shutdown system cost approximately $150 to $200 for any of the systems.

9.2.3 Inlet Fume Temperature Control

9.2.3.1 Recommended Practice

When a heat exchanger is installed to preheat a widely varying flow of inlet fumes with the combustion products, temperature control on the inlet fumes should be considered. The temperature can be controlled by either bypassing the fumes around the exchanger or by adding cool air to the fume stream ahead of the exchanger. Bypassing the inlet stream results in the most economical use of fuel but the heat exchangers must be constructed to withstand the temperature of the exhaust gases with reduced cooling from the inlet stream. Diluting the fumes with cool air provides continuous cooling for the heat exchanger but requires fuel to heat the added air up to the required oxidation zone temperature. Bypassing the hot gases around the heat exchanger is another approach but the bypass valve must be able to handle the hot exhaust gases. The selection of the temperature control method will depend upon an economic balance of increased fuel costs versus the increased construction costs.

If prolonged operation at reduced rates is planned, a control system should be provided that will continuously measure the inlet temperature and either add air to the inlet stream or bypass gases around the heat exchanger. If only occasional operation at reduced flow rates is expected, a simpler intermittent control system can be used. This system opens a damper to admit a given amount of air to the inlet stream whenever the temperature exceeds the safe value.

9.2.3.2 Equipment - Temperature Measurement and Control, Temperature Switches

Temperature Measurement and Control

The same type of equipment can be used here as was used for measuring and controlling the oxidation zone. However, since the temperature of the inlet stream is considerably lower than the oxidation zone, a gas or liquid filled transmitter can be used with the pneumatic or electronic process controllers if either of these systems are selected. The filled transmitters cost about $200 less than the thermocouple transmitters but a pneumatic or electronic control system will still be more expensive than the electric temperature controller operating directly from a thermocouple.

Temperature Switches

Thermocouple switches can be used here as well as in the combustion zone. Temperature switches based on the expansion of liquid in a bulb system are also satisfactory at a somewhat lower cost.

9.2.3.3 Typical Systems

Figure 9-4 shows a typical system where continuous control is maintained. The output from the controller goes to either the air dilution valve or the bypass valve but not both. The system shown in Figure 9-5 is typical for an intermittent control system.

Figures 9-4 and 9-5 follow

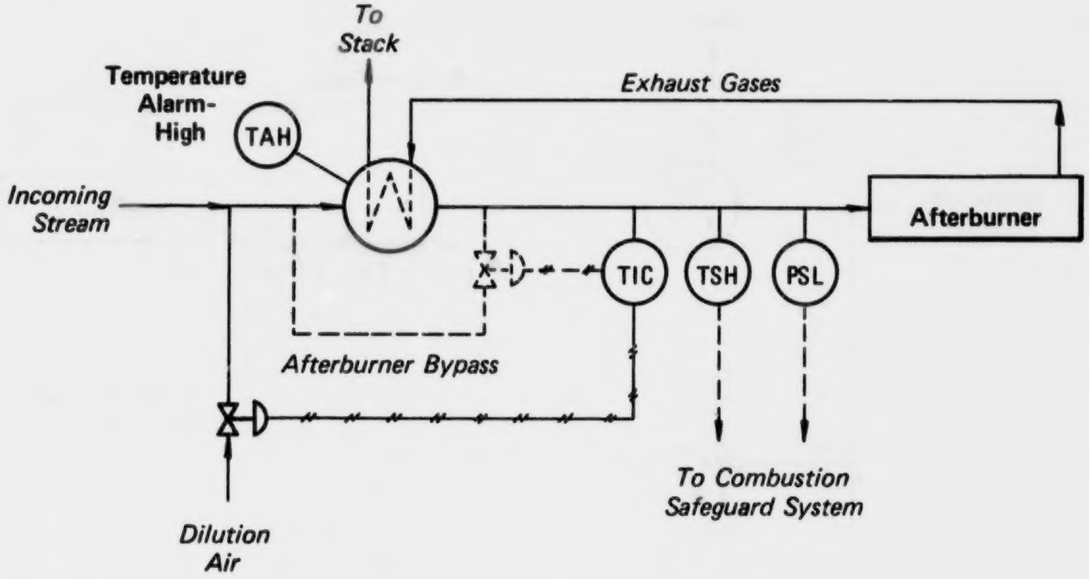

To Stack
Temperature Alarm-High
TAH
Exhaust Gases
Incoming Stream
Afterburner
TIC
TSH
PSL
Afterburner Bypass
To Combustion Safeguard System
Dilution Air

Figure 9-4. TYPICAL SYSTEM FOR CONTINUOUS CONTROL OF INLET GAS STREAM

136b

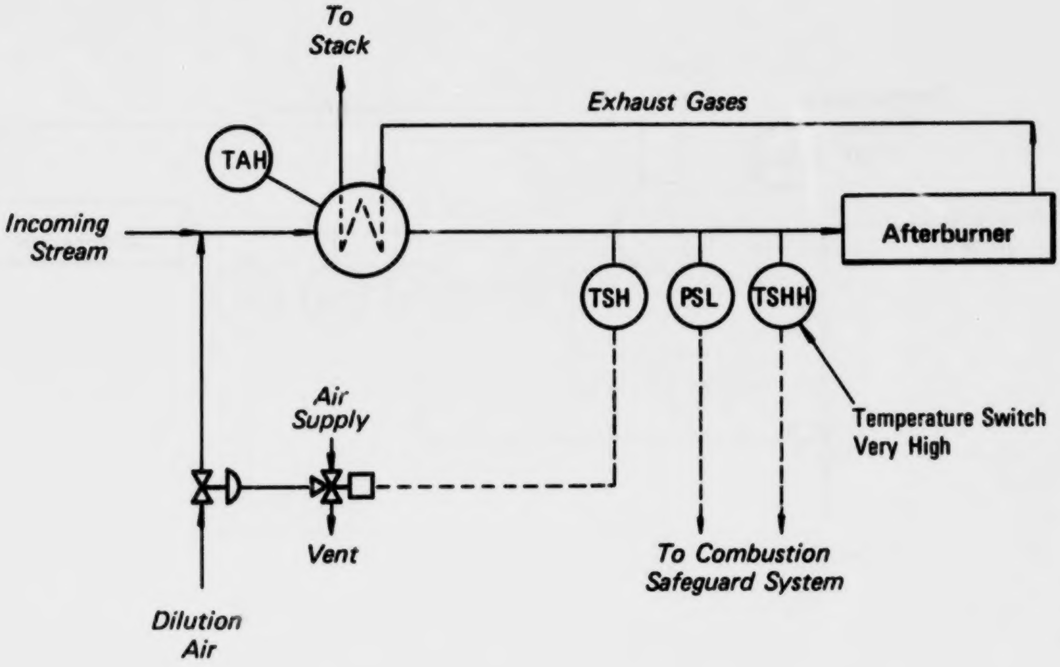

To Stack
Exhaust Gases
TAH
Incoming Stream
Afterburner
TSH
PSL
TSHH
Air Supply
Temperature Switch Very High
Vent
To Combustion Safeguard System
Dilution Air

Figure 9-5. TYPICAL SYSTEM FOR INTERMITTENT CONTROL OF INLET GAS STREAM

136c

9.3 Handling of Flammable Vapors

9.3.1 Recommended Practice

Detailed discussions of ventilation systems which handle flammable vapors are contained in the National Fire Codes published by the National Fire Protection Association and in the Handbook of Industrial Loss Prevention published by the Factory Mutual System. These agencies recommend that the flammable vapors be diluted with fresh air so that the concentration does not exceed 25% of the lower explosive limit at the maximum evolution of flammable vapors. This is especially important when a source of ignition such as an afterburner is located at the end of the ventilation system.

Interlocks must be provided that will shutdown the afterburner and vent the fumes in the event of failure of the ventilation fans. An air switch alone is not an acceptable method of interlocking the fans in ventilation systems where flammable volatiles are handled because condensation or deposits can cause the switch to become inoperative. Acceptable interlocking methods include rotational switches on the fan shaft, extra contacts on the fan starter, relay contacts operated in parallel with the fan motor, and powering the combustion safety systems from the load side of the fan starter. This insures that the fan is turning when the afterburner is operating.

There should be an automatically timed purge of the duct system with fresh air prior to lighting the afterburner to remove any flammable vapors or to evaporate any condensed material from the duct.

A vapor concentration indicator on streams with flammable vapors is highly recommended regardless of the concentration of fumes. This instrument, if properly installed and maintained, will give a continuous reading of the concentration in terms of the per cent of the lower explosive limit. In addition to indicating or recording the concentration to insure that safe conditions are maintained, by adding alarm or control features, the instrument can be made to control the vapor concentration by opening or closing dampers or to alarm and/or shutdown the afterburner if the vapor concentration exceeds a safe value.

When a continuous vapor concentration indicator or controller is installed and properly maintained, the concentration of flammable vapor in the inlet stream can be increased up to 50% of the lower explosive limit. Any increase in the concentration above this point should sound an alarm and shutdown the afterburner.

The use of vapor concentration controls on the dilution air can result in considerable saving in fuel cost when the afterburner is used in service where the vapor load varies. However, the measuring instrument and the controller must be properly maintained and calibrated or a dangerous condition can develop. This may require daily servicing in certain applications. If there is any doubt that adequate maintenance can be provided, sufficient dilution air should be provided to keep the vapor concentration below 25% of the lower explosive limit at all times.

In some special cases, flammable vapors may be generated with little or no air present. In this case, the mixture would be above the upper explosive limit and would not burn. Because of the chance of air leaking into a duct and creating a hazard, the usual practice is to dilute the vapors below the lower explosive limit at the source and to transport the diluted vapors to the afterburner. However, if it can be assured that the vapors will always be above the upper explosive limit and the local authorities and/or fire underwriters agree, the vapors could be treated as a fuel source, mixed with air at the afterburner and burned. A system such as this would require additional safeguards such as flame arrestors and blow out discs and possibly storage facilities to even out the vapor flow. Any system which handles vapors that are or can be made explosive or combustible by diluting with air should be approved by local authorities and the fire underwriters.

9.3.2 Equipment

9.3.2.1 Vapor Concentration Indicator

Most currently used vapor concentration detectors operate on the catalytic combustion principle. When a specially treated platinum filament is heated and contacted with a mixture of air and a flammable gas or vapor, combustion takes place at a temperature considerably lower than the normal ignition temperature. The heat generated by the combustion is sensed either by a thermocouple attached to the filament or by changes in the resistance of the filament itself.

The cell life of the detector measuring the resistance of the filament is shorter and more subject to zero drift than the model using thermocouples for temperature measurement. Therefore, the thermocouple type detector should be preferred to the type measuring the resistance changes of the filament itself.

The sample system for the detector is a very important item. The system must be constructed so that vapors do not condense before they reach the detector and the sample flow must be great enough to give a rapid response. The location of the sampling point must be selected to obtain a representative sample. A well mixed location is necessary since most flammable vapors are heavier than air and may stratify if the duct velocity is low.

A diffusion head analyzer is slightly different in the method of sample handling. The diffusion head is located at the sample point and does not require a sample pump or flow control system. Although large amounts of particulate matter, moisture and dust can cause plugging, the diffusion head should be used in clean, dry streams because of its simplicity and reliability.

To insure the safe operation of vapor concentration detectors, alarms should be considered to warn in case of:

Filament failure
Power failure
Low sample flow (not available for diffusion heads)

Figure 9-6 follows

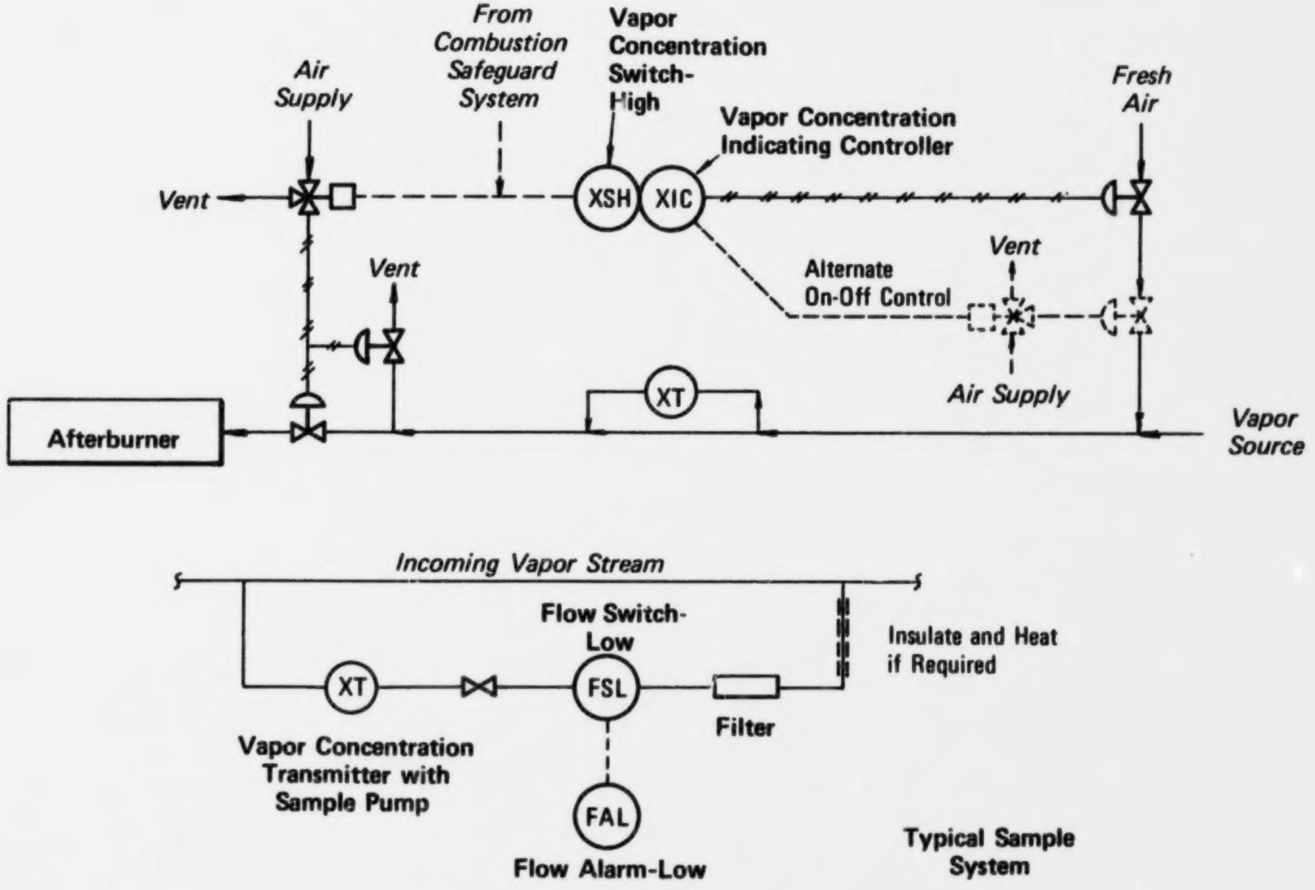

Air Supply
From Combustion Safeguard System
Vapor Concentration Switch-High
Vapor Concentration Indicating Controller
Fresh Air
Vent
XSH
XIC
Vent
Alternate On-Off Control
Vent
Air Supply
XT
Afterburner
Vapor Source
Incoming Vapor Stream
Flow Switch-Low
Insulate and Heat if Required
XT
FSL
Filter
Vapor Concentration Transmitter with Sample Pump
FAL
Flow Alarm-Low
Typical Sample System

Figure 9-6. TYPICAL VAPOR CONCENTRATION CONTROL SYSTEM

The cost of a vapor concentration monitoring system with sample system is between $1500 and $2000. The addition of control features can add another $1000.

9.3.2.2 Using the Afterburner Temperature Rise for Control

Split range temperature control is used to first turn down fuel to the burner as the fume concentration rises to some limit, e.g., 25%-50% of the LEL. At this point the fume stream will account for a certain fraction of the ΔT across the afterburner (~600°F). As fume concentration rises further, the controller operates a damper to provide more air flow through the oven or other source of fumes and prevents further concentration increase.

9.3.3 Typical Vapor Concentration Control System

The usual method of controlling the concentration of flammable vapors is by means of a damper to admit air to the duct. This can be a continuously adjustable damper if the fume load varies widely or a two position damper if the load swings between two relatively constant flow rates such as in a batch process. Figure 9-6 shows a typical arrangement. In addition to the control features there should also be alarms and shutdowns to protect the system from fan failure or high vapor concentration. In the event of a shutdown, the fumes should be bypassed around the afterburner.

9.4 Handling of Toxic Materials

9.4.1 Recommended Practice

All of the previous recommendations for handling flammable vapors should be followed where applicable. However, when the material is toxic and cannot be vented temporarily, additional precautions are required. Among the precautions to be considered are additional and/or redundant alarms to warn of possible trouble, a stand-by afterburner with automatic startup and switching, and temporary storage to hold the toxic material until it can be burned. Emergency shutdown of the fume generating process should also be provided. Particularly in special cases such as this, the local authorities having jurisdiction and the insurance underwriters should be consulted.

Chapter 10. MEASURING AFTERBURNER PERFORMANCE-ANALYTICAL METHODS

10.1 Introduction

In order to assess performance of an afterburner, concentrations of critical components at the entrance and exit of the afterburner must be determined. Typically these are total hydrocarbons and carbon monoxide but may include specific hydrocarbons, halogens, particulates, etc. As a bare minimum the afterburner will have to meet certain standards imposed by the local air pollution control agency, either emission standards for specific pollutants or a percentage destruction in passing through the afterburner. Other standards may prohibit any odor or visible plume. The operator of the fume incinerator will often be responsible for demonstrating compliance with regulations, either by doing his own testing or by hiring an outside laboratory.

If the afterburner does not provide the pollutant destruction required, some investigation into interior performance of the unit must be made. Of primary concern, for reasons discussed in Chapter 3, is measurement of two parameters: 1) Temperature - Is the temperature experienced by the fume stream high enough for the time the waste stream is in the afterburner? and 2) Flow Patterns - Has good mixing been achieved? Is there bypassing or severe non-uniformity of flows in the afterburner?

Certain measurements are also required before an afterburner is purchased. It is important that fume stream flowrate and composition be determined, both average values and the maximum variation. The amount and nature of entrained material will also influence design of the unit for reasons previously discussed.

In this Chapter an attempt is made to indicate the state of the art pertaining to the determination of a variety of air pollutants, measurement of temperature and determining flow patterns. Problems encountered in sampling and calibration are discussed. The methods available for determination of individual compounds or classes are summarized. Table 10-1 shows the components and concentration ranges which will be considered in this handbook. The high limits on these ranges are estimates of entering fume concentrations experienced in afterburner applications. The low values are based on estimates of stringent emission standards which may have to be met. Considerably more material is given for NO_x and SO_2 analysis than for other pollutants. This is an indication of the emphasis which has been placed on these components during the past decade.

In order to keep this Chapter to a reasonable size, no detailed procedures are included. References are given to sources of methods and appear in the Bibliography. The following references are particularly recommended to give an overall picture of the subject:

Jacobs10-25) and Leithe10-26) have given good reviews covering the subject of air pollutant analysis in general. Cooper and Rossano10-19) present a useful reference book covering sampling and analysis for particulate matter. Hanson et al10-12) give the most complete and detailed procedures covering wet

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chemical methods for determination of organic compounds in air. Ruch[10-27] reviews more compounds than Hanson, but the complete procedures are not given. Reference 10-45 is a good basic reference on temperature and flow measurement.

Odorous emissions are generally complex mixtures at very low concentrations which makes evaluation by chemical or instrumental methods difficult. Because of this, the usual approach to the measurement of odor emissions is to determine the magnitude of dilution required to obtain a concentration at which human judges can just detect the odor (the odor threshold). The strength of an odor is described in "odor units" and one odor unit is defined as the amount of odor necessary to contaminate 1 cubic foot of odor-free air to the threshold level. Odor measurements are at best semi-quantitative, but they have been useful in judging the effects of odor control measures (principally afterburners). It has been estimated[10-30] that an odor strength of less than 150 odor units per scf in stack gases will not cause community odor problems. It also appears that an emission of up to one million odor units per minute from a plant will not cause an odor nuisance.

10.2 Sampling Considerations

The collection of representative and meaningful samples is the most basic aspect of all practical analysis. Samples must accurately represent the conditions at the point of sampling and they must be preserved unchanged until the analysis is made. Sample size and sampling frequency are important considerations. Wherever possible, a continuous in-place analysis is preferable.

The flow rate, temperature, and pressure of the gas stream at the point of sampling must also be measured for calculation of mass flow of the components measured.

Sampling probes must be positioned with the above considerations in mind. Steam-tracing or heating-tape-tracing of sample lines may be necessary to prevent condensation of components at ambient temperature. Detailed recommendations for sampling atmospheres for analysis of gases and vapors is given in ASTM D 1605-60[10-1]. A method for sampling stacks for particulate matter is ASTM D 2928-71[10-1]. In the sampling of streams containing liquid or solid particulate matter, sampling for these particles is particularly critical owing to segregation according to particle size due to acceleration effects. Sampling in this case should approach iso-kinetic conditions, i.e., the sample stream at the sampling point and the main stream should be the same in linear flow rate and direction. An excellent discussion of the errors associated with anisokinetic sampling is given by Lundgren and Calvert[10-2]. Sampling is discussed further under 10.4.8 Particulate Matter.

10.3 Calibration of Instruments

Accurate calibration of monitors, especially in field use, presents difficulties as with any gas measuring devices. The preparation of meaningful standards has not been easy; a real advantage has become available through the

Table 10-1. COMPONENTS OF INTEREST FOR MEASURING AFTERBURNER PERFORMANCE

Components	Concentration Range, ppm
Hydrocarbons	
Total	10 - 20,000
Non-Methane	10 - 20,000
Paraffins	10 - 20,000
Olefins	10 - 20,000
Aromatics	10 - 20,000
Oxygenated Organics	
CO_2	0.5 - 10%[a]
CO	10 - 20,000
Acids	1 - 500
Aldehydes	1 - 1000
Particulates	
Combustible	0.01 - 10 grains/cu ft
Non-combustible	
Visible Emissions	
Other Components	
SO_2	5 - 20,000
SO_3	5 - 1,000
H_2S, Mercaptans	1 - 10,000
Halogens	5 - 10,000
NO_x	5 - 5,000
Water	0.1 - 10%[a]
Oxygen	0.1 - 20%[a]

a) Not a pollutant, but determination may be needed for control.

use of permeation tubes[10-17,23]. These are Teflon tubes which contain the liquefied substance for which the calibration is to be made. Tubes are available for sulfur dioxide, nitrogen dioxide, hydrogen sulfide, chlorine, hydrogen fluoride, ammonia, carbon disulfide, carbon tetrachloride, acetaldehyde, several mercaptans and hydrocarbons. Tubes can be made to order for many compounds boiling above -70°F[10-18]. Each will generate a low concentration of the substance as a function of temperature.

A calibrated permeation tube containing the desired liquefied gas is maintained at a constant temperature while air flows over the tube at a fixed rate. From the permeation rate at the fixed temperature and the flow rate of the diluent gas the concentration of the test gas can be calculated. This mixture of known concentration is used to calibrate the analyzer. A typical calibration procedure as applied to sulfur dioxide is described in ASTM D 2914[10-1].

Several manufacturers (see Table 10-8) sell calibration equipment for use with permeation tubes.

10.4 Principles Involved in Analyses

The components of interest are discussed separately below. Under each component, a brief discussion of the chemistry involved in the existing analytical methods is given. Commercial instruments which have been developed are listed. Wherever a choice of several approaches is available, an attempt is made to recommend the optimum one for a given concentration range.

10.4.1 Hydrocarbons

Hydrocarbons in low concentration are measured by a flame ionization detector (FID). The principle of the FID is based on the linear relationship of ion formation to the concentration of a given organic compound in the flame. The gaseous mixture containing hydrogen is burned at a small jet with air or oxygen and the change in electrical conductivity is measured. Response is a function of the number and type of oxidizable carbon atoms in a molecule. The relative response of a carbon atom follows the decreasing order $-CH_2-$, $-CHOH-$, $-CO-$. Relative response factors of some typical compounds are shown in Table 10-2. In general, there is a small variation in the response of widely differing hydrocarbon types.

Oxygenated organic compounds have a lower relative response, decreasing with increasing oxygen content. CO_2 and CO, and inorganic gases show essentially no response. Chlorinated hydrocarbons also are measured; response is dependent on the number of CH bonds in the compound.

Federal regulations on ambient air quality standards are concerned with "non-methane" hydrocarbons rather than total hydrocarbons. This is due to the fact that in some areas appreciable concentrations of methane exist from natural sources and furthermore, methane does not become involved in photochemical smog reactions so it is not a problem.

Table 10-2. RELATIVE REPONSE FACTORS FOR THE FLAME IONIZATION DETECTOR[10-3,4a]

Compound or Type	Relative Response Factor per gram (n-heptane = 1.00)
Paraffinic hydrocarbons	
methane	0.97
propane	0.98
n-pentane	1.04
n-octane	0.97
2,3,4-trimethylpentane	0.99
Cyclic hydrocarbons	
cyclopentane	1.04
1,1 dimethylcyclopentane	1.03
ethylcyclohexane	1.01
Olefinic hydrocarbons	
acetylene	1.07
ethylene	1.02
hexene-1	0.99
Aromatic hydrocarbons	
benzene	1.12
p-xylene	1.00
tert-butylbenzene	1.02
Oxygenated compounds	
methanol	0.23
n-propyl alcohol	0.61
octyl alcohol	0.85
formaldehyde	0.1
acetaldehyde	0.3
acetone	0.49
diethyl ether	0.58
n-propyl acetate	0.53
formic acid	0.01
acetic acid	0.24
hexanoic acid	0.63
Chlorinated compounds	
chloroform	0.082
carbon tetrachloride	0.045
1,3 dichloropropane	0.37
trichloroethylene	0.23
tetrachloroethylene	0.19
Inorganic Gases	
argon	essentially no response
nitrogen	
oxygen	
carbon monoxide	
carbon dioxide	
water	

Instruments previously available measured total hydrocarbons including methane. A new group of instruments are now becoming available which will measure both total hydrocarbons and methane separately, thus giving the desired non-methane hydrocarbon value by difference. The technique can also determine carbon monoxide separately following its catalytic hydrogenation to methane. A measured volume of air is sampled and the total hydrocarbon content is measured with a flame ionization detector. An aliquot of the same air sample is passed through a stripper column which retards all hydrocarbons other than methane. The eluted methane is then measured by the flame ionization detector. The stripper column is then automatically backflushed to prepare it for subsequent samples on a semi-continuous basis.

The Los Angeles Air Pollution Control District describes analytical methods for organic solvents and vapors10-24). Total organic materials are measured by a total combustion analyzer (TCA). In this equipment, a large sample is passed through a dry-ice trap. The trap is then warmed through 2 temperature levels so that low-boiling and later, high-boiling components are evaporated into a carrier gas stream consisting of 5% oxygen in nitrogen. The gas mixture then passes through a catalytic oxidizer where organic compounds are converted to CO_2, which is measured in an infrared analyzer. Two peaks for organics result - one for low-boiling and the other for high-boiling materials.

Individual hydrocarbons are determined by gas chromatography using a flame ionization detector. A typical method for determination of C_1 to C_5 hydrocarbons is described in ASTM D 282010-1). The lower limit of measurement of the individual components is in the parts per billion range.

10.4.2 Oxygenated Organic Compounds, Carbon Monoxide, Carbon Dioxide

Since oxygenated organic compounds respond to the FID and are measured as hydrocarbons, they may be included in the category of hydrocarbons. However, in most instances, the oxygenates can be determined specifically by virtue of their functional groups. These may be determined by absorption in suitable reagents followed by a titration, colorimetric, or spectroscopic measurement. For example, carboxylic acids are readily absorbed in basic solution and determined by titration. Aldehydes and ketones react with 2,4-dinitrophenylhydrazine to form a highly colored product which can be measured spectrophotometrically. Table 10-3 lists a number of possible approaches which may be considered for several oxygenated species which may be encountered.

10.4.3 Chlorinated Hydrocarbons, Phosgene, Hydrochloric Acid

The use of the flame ionization detector for measurement of chlorinated hydrocarbons has been cited under "Hydrocarbons". In addition, other procedures are available for determination of carbon tetrachloride, chlorobenzene, chloroform, ethylene dichloride, tetrachloroethane, trichloroethylene, and ethylene chlorohydrin. These procedures involve absorption in suitable solvents, color development, and measurement of the color by spectrophotometric means.

Table 10-3. METHODS FOR DETERMINATION OF OXYGENATES AND OTHER COMPONENTS

Compound Determined	Method	Lower Limit, ppb	Absorptivity, L/mol cm	Reference
Aldehydes	Methylbenzothiazolone Hydrazone Spectrophotometric (650 nm)	20	50,000	4b,5,8
Aldehydes and Ketones	Dinitrophenylhydrazine Spectrophotometric	40	27,000	4c, 6
Acrolein	4-Hexylresorcinol Spectrophotometric (605 nm)	20	17,000	7, 8
Formaldehyde	Chromotropic Acid Spectrophotometric (570 nm)	20	19,000	4d, 8
Carboxylic Acids	Absorption-titration			9
Esters	Hydroxamic Acid Spectrophotometric (530 nm)	100	1,100	4e, 10
Carbon Monoxide	Non-dispersive Infra-Red (NDIR) (see Table 10-4)	1000		
Carbon Dioxide	NDIR (see Table 10-4) Orsat.	2000 0-5%		11

Carbon tetrachloride, chloroform, ethylene dichloride, trichloroethylene, tetrachloroethylene and ethylene chlorohydrin are all determined by a method involving absorption in pyridine, reaction with aqueous caustic, and measurement of the resulting color[10-12a]. The intensity of the color varies with the compound and the method should be calibrated against known quantities of the particular compound being measured. In general the method is sensitive in the range 10 - 1000 ng of the compound.

Chlorobenzene is measured by a method which involves absorption and reaction with a solution of formaldehyde in concentrated sulfuric acid. The yellow color produced is measured. The reaction is not specific for chlorobenzene, as aromatic hydrocarbons in general are measured[10-12b].

Upon passing through an afterburner, chlorine containing compounds should produce hydrochloric acid as a major product and traces of phosgene. The method most generally applied for hydrochloric acid involves absorption in aqueous caustic, acidification and titration with silver nitrate[11-12c]. Phosgene is determined by a number of colorimetric procedures[10-27]. In a typical procedure[10-25], the air sample is drawn through a filter impregnated with ethylhydroxyethylaniline and dimethylaminobenzaldehyde. Phosgene produces a blue color whose intensity is proportional to concentration.

10.4.4 Oxygen and Nitrogen

Oxygen and nitrogen content may be of interest for the purpose of calculating excess air in the process. An Orsat analysis can be applied to measure CO_2 and oxygen. Nitrogen is determined as the difference between 100 and the sum of the CO_2 and oxygen. The concentrations are in the 2 - 20% range, where the Orsat equipment is of adequate accuracy, (see EPA method 2)[10-11]. Process analyzers are also available for oxygen and CO_2.

Table 10-4. ANALYZERS FOR O_2 AND CO_2

Component	Instrument	Principle
O_2	Beckman C2 Beckman Model 715 Teledyne 320P	Paramagnetic Electrochemical sensor
CO_2	Beckman IR-215 variety of concentration ranges is available.	Non-dispersive Infra Red (NDIR)

10.4.5 Nitrogen Oxides (NO_x)

A number of different oxides of nitrogen exist but only three are of importance in the atmosphere. These are

N_2O	Nitrous Oxide
NO	Nitric Oxide
NO_2	Nitrogen Dioxide

Since nitrous oxide is not toxic and it does not appear to take a part in photochemical smog reactions, its determination is not of general concern in air pollution problems.

NO is produced in appreciable amounts in combustion processes and whenever air is heated to a high temperature. The reaction

$$N_2 + O_2 \rightleftharpoons 2\,NO$$

has an equilibrium constant which increases markedly at elevated temperatures. The nitrogen oxides from high-temperature sources are predominantly (98 - 99%) in the form of NO. In the presence of excess oxygen, NO reacts at a moderate rate to produce NO_2, which is in equilibrium with the tetroxide (N_2O_4). At the low concentrations at which the oxides of nitrogen exist in the normal atmosphere the equilibrium

$$N_2O_4 \rightleftharpoons 2\,NO_2$$

is almost entirely on the side of the dissociated species. Of the oxides of nitrogen, N_2O and NO are colorless, while the monomeric NO_2 is deep brown.

The concern regarding NO and NO_2 in air pollution involves three areas:

i These oxides of nitrogen are chain carriers in the generally accepted mechanism for photochemical smog formation,

ii The brown tinge of the Los Angeles atmosphere may be due to the NO_2 present, and

iii Limited tests indicate that NO_2 may be toxic at parts per million levels.

The applicability and limitations of the principal methods for the determination of NO_x are outlined in Table 10-5. The two chemical methods have been used for some years and, in general give good agreement in the range of 10 to 4000 ppm. That involving the reaction of phenol disulfonic acid to a nitrated product (ASTM D 1608)$^{10-1)}$ is of moderate sensitivity and can be employed for the determination of oxides of nitrogen, with the exception of nitrous oxide, at levels above 5 ppm. This method is generally applied to samples taken in

Table 10-5. SUMMARY OF APPLICABLE METHODS FOR NO_x

METHOD	PRINCIPLE OF PROCEDURE	INSTRUMENTATION	APPLICABILITY	LIMITATIONS
Phenol disulfonic acid method ASTM D 1608 (Method 4 EPA)	$NO_x \xrightarrow[H_2SO_4]{H_2O_2} HNO_3$ Phenol disulfonic acid nitrated to produce yellow color.	Laboratory spectrophotometer at 400 mμ	Range: 5 to 1000 ppm for all nitrogen oxides except N_2O.	Not sensitive below 5 ppm.
Sulfanilic Acid method (Saltzman) ASTM D 2012 SAE J177	NO_2 converts sulfanilic acid to diazonium salt. Salt couples with amine to produce deep violet color	Laboratory spectrophotometers at 550 mμ. Continuous analyzers: Beckman Acralyzer. Technicon Autoanalyzer.	Range: 0.01 to 4000 ppm. Specific for NO_2 NO determined by prior oxidation. Useful for air and exhaust analysis. Faster than disulfonic acid method	Equipment somewhat more complex than for phenol disulfonic acid method
Ultraviolet absorption method	NO_2 has maximum at 400 mμ with absorptivity = 170 liters/mole-cm. NO is transparent above 230 mμ.	Continuous Analyzers: Beckman NDUV Model 255 plus oxidation system.	Range: 10 to 6000 ppm. Determines NO_2 directly. NO determined by prior oxidation.	NO is a reactive gas and can be partially lost unless precautions are taken.
Infrared absorption method	NO has band at 5.4 μ with absorptivity = 2 liters/mole-cm.	Continuous Analyzers: Beckman NDIR Model 315A. Mine Safety Appliance Model LIRA 200.	Range: 10 to 4000 ppm. Determines NO directly.	Water vapor interferes and must either be constant, or preferably removed.
Electrochemical Sensor	NO+NO_2 permeate membrane on sensor and are electrochemically oxidized. Resulting current is proportional to NO_x concentration.	Continuous Analyzers: Dynasciences NX-110 and NX-130 EnviroMetrics Model N-122 Theta Sensors Model LS-800-ANX	Range: 2 to 10000 ppm Models available for NO_x or NO_2	SO_2 interferes but can be eliminated or compensated for.
Chemiluminescence	The light resulting from the reaction of NO with ozone is measured with a photomultiplier. NO_2 must be converted to NO to be measured.	Continuous Analyzers: Thermo-Electron Corp. Bendix, Environmental Science Division REM, Inc. Model 642	Range: 0.01 to 10000 ppm	No known interferences

flasks which contain an oxidizing absorbent. The phenol disulfonic acid method is used by a number of local control districts including the Los Angeles County Air Pollution Control District[10-13,14].

The sulfanilic acid method, otherwise known as the Saltzman or Griess-Saltzman method, is of considerably higher sensitivity than the phenol disulfonic acid methods and is specific for nitrogen dioxide. It must be stressed that NO, the predominant species present in incinerator effluent, is not determined directly by the sulfanilic acid reaction and it must first be oxidized to NO_2 for its determination. An SAE method for NO_x in diesel exhaust[10-15] involves taking a sample into a syringe along with air and sulfanilic acid reagent. The NO is oxidized by the air during a reaction period and the color formed is a measure of the NO_2 plus NO originally present. ASTM D 2012[10-1], a method designed for ambient concentrations, measures NO after oxidation to NO_2 by permanganate.

Spectrophotometric instrumentation is particularly useful for the monitoring or analysis of the nitrogen oxides in combustion exhausts. The infrared region is suitable for the determination of NO and this can be done effectively with infrared instrumentation using a narrow band-pass non-dispersive filter to provide discrimination against interfering components. Either the Beckman NDIR 315A or the Mine Safety Appliance LIRA 200 offer feasible direct measurement of NO in the range of 10 to 4000 ppm with minor interference except for water vapor which must be kept constant or preferably eliminated.

In the ultraviolet and visible region, NO is transparent above 230 mμ but NO_2 has a strongly absorbing band with a maximum in the vicinity of 400 mμ. The Beckman Model 255 Nitrogen Dioxide Analyzer is a nondispersive ultraviolet (NDUV) spectrophotometer and is effective for the direct determination of NO_2 in effluents. For the application of the ultraviolet technique to the determination of NO or total nitrogen oxides, it is necessary to convert the NO into NO_2 by oxidation before the gas sample is introduced into the sample cell. The Beckman Model 255 (NDUV) Nitrogen Dioxide Analyzer, when used in conjunction with the Beckman Model 315A (NDIR) Nitric Oxide Analyzer offers a direct, continuous system for the simultaneous measurement of NO and NO_2. For the best results both water vapor and hydrocarbons should be removed and this can be done with cold traps and Drierite dehydrating agent without excessive loss of the nitrogen oxides.

Instrumentation for measuring nitrogen oxides by means of electrochemical sensors has been available since 1969. The sensor consists of a pair of electrodes immersed in an electrolyte. A permeable membrane covers one electrode and encloses the cell. The NO_x ($NO+NO_2$) present in the sample gas can permeate the membrane and be electrochemically oxidized at the applied potential. The resulting current, which is proportional to the NO and NO_2 present, is amplified and read out on a meter or recorder. The NO_x sensor also responds to SO_2 on an approximately equivalent basis. The SO_2 interference can be removed by using a selective scrubber solution provided by Dynasciences, or by electrical compensation provided by a second sensor for SO_2 with the EnviroMetrics and Theta Sensors instruments. Carbon monoxide can interfere to a slight extent if present in high amounts (about 1% or more).

The chemiluminescence method for NO is quite new but appears to have great promise. It is based on the reaction between NO and ozone which results in the emission of light. The light is measured with a photomultiplier tube. The method is very sensitive and can be used to determine low ambient levels of nitrogen oxides. However, NO_2 must first be converted to NO by passing the sample gas through a furnace unit. At present there are no known interferences to the method.

10.4.6 Sulfur Dioxide

Since regulations limit the maximum concentrations of SO_2 in air to the range of tenths of a part per million, sensitive and reliable analytical methods and monitoring instruments have been developed for this determination. Some of the important features of the principal methods in current use are summarized in Table 6. Although there are a number of variations of these methods, they are essentially based on four types of chemical reactions.

10.4.6.1 Hydrogen Peroxide Methods

SO_2 in intimate contact with dilute hydrogen peroxide solution is dissolved and oxidized rapidly according to the reaction:

$$SO_2 + H_2O_2 \longrightarrow H_2SO_4$$

The resultant sulfuric acid can be titrated with base or the sulfate determined by titration with barium perchlorate. These titration techniques find most common use in the determination of the SO_2 concentration at higher levels than is normally encountered in the atmosphere, such as in stack gases at levels of 200 to 3000 ppm. This approach is probably most useful for incinerator effluent.

For the determination of low concentrations of SO_2 in air, advantage can be taken of the increase in conductivity of the hydrogen peroxide solution due to the formation of sulfuric acid which is highly dissociated. Automatic analyzers based on this principle have found use since they can provide continuous results in the range of 0.01 to 5 ppm with little need for operator attendance. The scrubber in the conductivity instrument is usually filled with a dilute hydrogen peroxide solution containing sulfuric acid in order to avoid interference from carbon dioxide. The conductivity instruments must be calibrated by procedures based on primary standards. It is important to realize that the conductivity method is not specific for SO_2 since other electrolyte-producing materials can interfere. In making measurements near the ocean, salt aerosols have been found to interfere; acidic and basic gases or aerosols which are likely contaminants in some industrial areas, are also common interferences. The useful range of this equipment, 0.01 to 5 ppm, may be too low for analysis of incinerator effluent.

10.4.6.2 Coulometric Methods

The reactions involved in the coulometric methods are essentially the following:

Table 10-6. METHODS FOR SO_2

Method	Principle	Instrumentation	Applicability	Limitations
Hydrogen Peroxide Titration Method (EPA method 3)	$SO_2 + H_2O_2 \rightarrow H_2SO_4$ Titration with base or Ba^{++}	Laboratory Equipment: Absorber and titration units	Range: 0.01 to 100 ppm	Requires reagent additions.
Hydrogen Peroxide Conductometric Method	$SO_2 + H_2O_2 \rightarrow H_2SO_4$ Measure Conductivity	Monitors: Leeds & Northrup, AEROSCAN Wosthoff U3S ULTRAGAS ANALYZER Instruments Development, IDC 902-I Scientific Instruments, SI-67	Range: 0.01 to 5ppm	Interference by salt aerosols and acidic and basic gases which may be eliminated by filters.
Electrolytic (Coulometric) Method	$SO_2 + Br_2 + H_2O \rightarrow 2HBr + H_2SO_4$ $SO_2 + I_2 + H_2O \rightarrow 2HI + H_2SO_4$ Br_2 and I_2 generated electrolytically.	Monitors: Consolidated, TITRILOG Beckman Instruments, Model 906 Barton 286 SULFUR TITRATOR Phillips Instruments, Model PW 700 Atlas Electric Devices, Model 1200	Range: 0.01 to 5 ppm Monitors simple to operate and reliable for unattended service.	Interference by oxidizing materials, aldehydes, olefins, and hydrogen sulfide. (Some interference can be eliminated by filters).
WEST-GAEKE Colorimetric Method ASTM D 2914	Formation of dyestuff by reaction with bleached pararosaniline.	Laboratory Equipment: Spectrophotometer Monitors: Atlas Electric Devices, Model 1500 Technicon Corp., AUTO ANALYZER	Range: 0.01 to 5 ppm Most nearly specific method for SO_2.	Procedure cumbersome. Continuous analyzer needs close attention.
Electrochemical Sensor	Oxidation in a membrane-covered cell	Monitors: Dynasciences, SS-330 Envirometrics NS-200 Theta Sensors LS-800-AS	Range: 0 to 5000 ppm Simple to operate	NO and NO_2 interfere slightly.

$$SO_2 + I_2 \text{ (or } Br_2\text{)} \longrightarrow 2HI \text{ (or } 2HBr\text{)} + H_2SO_4.$$

Instruments are commercially available which operate on the principle of coulometric generation of free halogen to react with the SO_2 removed by the cell from the gas stream. The current generated to produce sufficient halogen to satisfy the reaction above is a measure of the SO_2 content of the inlet gas stream. These coulometric instruments can be used in the range of 0.01 to 5 ppm of SO_2 in the atmosphere. Oxidizing substances such as chlorine, NO_2 and ozone will give low results for SO_2, but the oxidants may be removed in some cases by a filter containing ferrous sulfate crystals. Oxidizable compounds (other than SO_2) such as hydrogen sulfide, mercaptans or aldehydes, will give a positive error unless removed by a suitable filter system. Some of the commercial instruments based on the coulometric technique are finding extensive use as monitors since they are quite simple to operate and require only minor attention.

10.4.6.3 Colorimetric Method

The method generally considered most specific for the determination of SO_2 is that of West and Gaeke(10-16) which involves the initial fixation of SO_2 as a stable complex with sodium tetrachloromercurate. Addition of a mixture of acid-bleached pararosaniline and formaldehyde to the complex produces the deep red pararosaniline methyl sulfonic acid.

The detailed procedure published as ASTM D 2914(10-1) for conducting the West-Gaeke method is recommended for the determination of concentrations of SO_2 in the range of 0.003 to 5 ppm. Acidic or basic gases or solids (sulfur trioxide, sulfuric acid, ammonia, or calcium oxide) cause no interference with the method, but the gaseous oxides (ozone and NO_2) interfere if present in concentrations greater than that of the SO_2. The interference of NO_2 can be eliminated by adding a small quantity of sulfamic acid during the analysis. The interference due to ozone can be reduced by allowing the sample solution to stand, and the ozone to decompose, prior to analysis.

Although the West-Gaeke method, when employed with due precautions concerning possible interfering materials, is considered the most specific for the determination of low concentrations of SO_2 in the air, the procedure is elaborate and the continuous analyzers available are cumbersome. Since the method requires close attention, it is not as widely used as the conductivity or the coulometric techniques.

10.4.6.4 Electrochemical Sensor Method

The same instruments of this type used for determination of nitrogen oxides can be used for the measurement of SO_2 by interchanging sensors or by using a dual unit with sensors for both NO_x and SO_2. The same type of cell is used but the selectivity is altered by changing the applied potential, and perhaps the electrolyte and membrane. The current resulting from the electrochemical oxidation of the SO_2 is proportional to the SO_2 concentration. NO may give a slight positive interference while NO_2 may give a negative response. The extent of the interferences will vary some depending on manufacturer.

10.4.7 Sulfur Trioxide, Sulfuric Acid

Any combustion process in which a sulfur containing fuel is used will usually produce some sulfur trioxide along with the sulfur dioxide. The SO_3 may range from 0.5% or less up to about 10% of the total sulfur oxides emitted and the concentration may range from about 1 ppm up to 50 ppm or more.

Many methods have been published for the determination of SO_3 in the presence of SO_2. These usually consist of separating the SO_3 and the SO_2 and then determining the resulting acidity or sulfate ion from each. The two most widely used techniques for separating the SO_3 and SO_2 are based on absorption[10-14,43)] and condensation[10-44)]. In the one case, the sample gas is passed through a set of absorbers in which the SO_3 is first removed by an isopropyl alcohol-water solution and the SO_2 then removed by an aqueous hydrogen peroxide solution. Some SO_2 dissolves in the first IPA-H_2O scrubber solution and must be subsequently purged out with nitrogen to transfer it all into the peroxide scrubber. The IPA inhibits the oxidation of SO_2 to SO_3 before the separation is completed. In the second case, advantage is taken of the fact that as the hot combustion gases are cooled down, the SO_3 combines with water vapor and becomes a sulfuric acid mist. This sulfuric acid mist is then condensed out at a temperature below its dew point, but above the dew point of water vapor. The SO_3 is thereby separated from the SO_2 which passes through the condenser to a peroxide scrubber.

The solutions obtained following the separation of SO_3 and SO_2 may be titrated for acidity with standard base if no other acidic interferences are present. More frequently the resulting sulfate ion is determined by titration with barium ion using thorin indicator, or colorimetrically following reaction with barium chloranilate.

At present there are no commercially available instrumental methods for determining SO_3 and it must be determined by the manual wet chemical methods outlined above.

10.4.8 Particulate Matter

Particulate matter can be emitted from incinerators for at least three reasons: combustion is incomplete and soot is formed, some of the feed to the incinerator escapes combustion and appears as a condensible vapor in the stack gas, or the fume contains inorganics which appear as oxides. In any event, the effect of a significant amount of particulate matter in the incinerator stack gas is the formation of a visible plume. In most air pollution control districts, plumes above a certain opacity are illegal and steps must be taken to prevent their formation. In addition to compliance with control district regulations, there are sometimes other reasons for controlling particulate matter emissions. These include the use of soot formation as a guide to controlling the combustion process, and the prevention of the loss of economically valuable ash products. The first step in such control is the estimation of the amount of particulate matter in the stack gas. Cooper and Rossano[10-19)] review the methods for this purpose and cite 15 references to methods used by regulatory agencies.

The most easily applied test for the amount of particulate matter in a stack gas emission is the visual evaluation of the plume opacity. It has been found to be possible to train observers to estimate the plume opacity to within 10 per cent light transmission. The unit commonly used in this field is the Ringelmann Number. The appearance of the plume is compared with a series of standards prepared by ruling a white surface with sets of lines of increasing width. In the Ringelmann standard, 20 per cent of the surface is covered by black lines, in Ringelmann 2, 40 per cent, and so on up to Ringelmann 5 which is completely black.

In the basic method for the determination of Ringelmann Number, large cards ruled in the manner described are placed about 50 feet away from the viewer and in the same field of view as the plume. The lines then fuse into fields of grey of different density. The number assigned to the plume is the number of the standard which most closely matches the plume in appearance. The basic method is often impractical due to restricted access and the need for viewing the plume from several vantage points. Consequently, control districts often train their observers to evaluate the plume opacity from memory. Miniature cards are available as guides for occasional observers but are no substitute for training by the evaluation of standardized plumes. Such training becomes especially important in the evaluation of white or colored plumes.

The accurate sampling of incinerator stack gases for particulate matter content is a difficult job. The difficulties arise from the nature of the medium which is sampled. It is usually hot (over 1000°F), wet, and moving at a velocity of one to several thousand feet per minute. Furthermore, the composition of the gas sometimes varies from point to point in the duct and with time. The concentrations of interest are often low (.01 to .1 grains/cubic foot) and sometimes the gas contains condensible vapors other than water which form particles on cooling in the ambient air. For these reasons, there are no simple techniques which give reliable particulate matter concentrations in all circumstances. The methods used by air control districts generally require a day or so for a skilled two man team to test an installation thoroughly. The method outlined in the next paragraph is typical.

The apparatus used in the test is shown in Figure 10-1 and is the one described by Smith[10-20]. During the test, sample gas is withdrawn through the train at a rate such that the velocity of the gas through the nozzle matches that of the gas in the duct. In order to achieve this rate, the duct velocity is measured with the aid of the Pitot tube. Calculations made with the aid of a nomograph indicate the pressure drop to be established across the orifice meter. Particles existing as such at the stack conditions are collected in the cyclone and filter. Water and other condensible vapors are collected in the impingers. During a test, the nozzle is moved to several preselected locations in the duct to ensure the collection of a representative sample. At the end of the test, the particles deposited in the probe are combined with those in the cyclone and filter and weighed. The impinger contents are evaporated to dryness at about 220°F and weighed also. The particulate matter concentration is calculated from the sum of the weights of material, both in the dry collectors and probe and in the impingers.

Accurate measurement of particulate matter content is probably impractical for any purpose except compliance testing. However, less rigorous techniques can give information which is useful for control purposes. These techniques use monitors to provide automatic records of some property of the stack gas related to the particulate matter burden. Cooper and Rossano[10-19] discuss the state of the art here and point out that there are two main approaches to particle monitoring. The first employs the attenuation of the intensity of a light beam through the stack and the second employs tape samplers.

The elements of an in-stack smoke photometer are shown in Figure 10-2[10-21]. They consist of a light source and photocell mounted on opposite sides of the stack. The windows in the light path are kept clean by streams of fresh air directed over them. Output from the photocell is directed to a readout device which can be an indicator, recorder or alarm. This kind of unit is widely used in large power generating stations and other combustion installations. Because the light absorption depends upon particle size as well as concentration, the smoke photometer record is related more to Ringelmann number than to weight concentration of particulate matter. Moreover, vapors which would be counted as particulate matter in a wet collection method such as that described above, would not be seen by these photometers. However, these units do not require special sampling methods nor is the temperature or water content of the stack gas a problem.

Another monitoring technique uses the tape sampler. Gas is drawn through a filter paper tape at a fixed rate for a given time. Then the tape is automatically advanced to make the spot visible and another sample taken. Models of sampler are available which measure the light transmittance of the spot right after the tape advance and print out the value on paper tape. Tape sampling requires more attention to the means used to convey the stack gas to analysis point than do the photometers. A sample is withdrawn through a probe which must be located properly and at a rate which corresponds to the velocity of the stack gas if a representative sample is to be taken. When the gas is hot or wet, a known amount of dilution gas must be added to bring the mixture to ambient conditions without precipitating the water. Tape samplers with optical readers are subject to changes in output signal introduced by changes in the color of the particle or anything else which affects the optical density of the spot on the tape. The complications introduced by the sampling problem and the intermittent nature of the output make the tape sampler less desirable than the photometer for most installations.

A portable opacity meter is manufactured by Shell Development Company for the measurement of stack plumes. The method is based on using the sun as a light source and the measurement is expressed as Ringelmann number or as percent transmissions. The instrument is adjusted to read 100% transmission in un-obstructed sunlight. Then the observer places himself so that the plume is between him and the sun and a reading of the sun obscured by the stack plume is made. The light path of the instrument contains optical filters which produce an instrumental spectral response closely approximating that of the human eye. A silicon photodiode converts the light to an electrical signal which is amplified and displayed on an internal meter. Since the legal constraints placed on plume opacity are based on visual effects, any instrumental technique should measure the same quantities. The instrument can be used under most weather conditions, except when the sun is obscured or when moving clouds change the available sunlight during calibration or measurement. Under normal conditions the instrument's precision is ±0.25 Ringelmann number, markedly better than the trained human eye which usually cannot discriminate beyond ±0.50 Ringelmann number.

There are several good references available in the field of particulate matter sampling. The most complete and the most recently published of them is the work by Cooper and Rossano$^{10-19)}$. They provide a comprehensive guide to the literature, together with a good summary of the corpus of knowledge on this subject. Particularly valuable to the man in the field is the collection of numerical data and computation methods. Somewhat older are Chapter 28 and 29 in the work edited by Stern$^{10-9)}$. These cover much of the same ground as do Cooper and Rossano but with less depth and less practical information. The methods used by the Los Angeles Air Pollution Control District are described by Devorkin et al$^{10-22)}$. A recently published ASTM method, D2928-71$^{10-1)}$, describes methods for measuring particulate matter concentration.

10.4.9 Source Measurement of Odor Emissions

Odor is a physiological response of individuals, and it is because of this response that odor complaints are made. There is no instrument that parallels the nose in the observation of odors and because of this all measurements of odor emission ultimately rely on observations made by the human nose.

Odorous emissions are generally complex mixtures at very low concentrations which makes evaluation by chemical or instrumental methods very difficult. Because of this the usual approach to the measurement of odor emissions is to measure odor intensity by determining the magnitude of dilutions required to obtain a concentration at which human judges can just detect the odor (the odor threshold). The techniques to accomplish this will make up the main part of the discussion which follows. Some note will also be made of progress toward physico-chemical techniques for the

Figures 10-1 and 10-2 follow

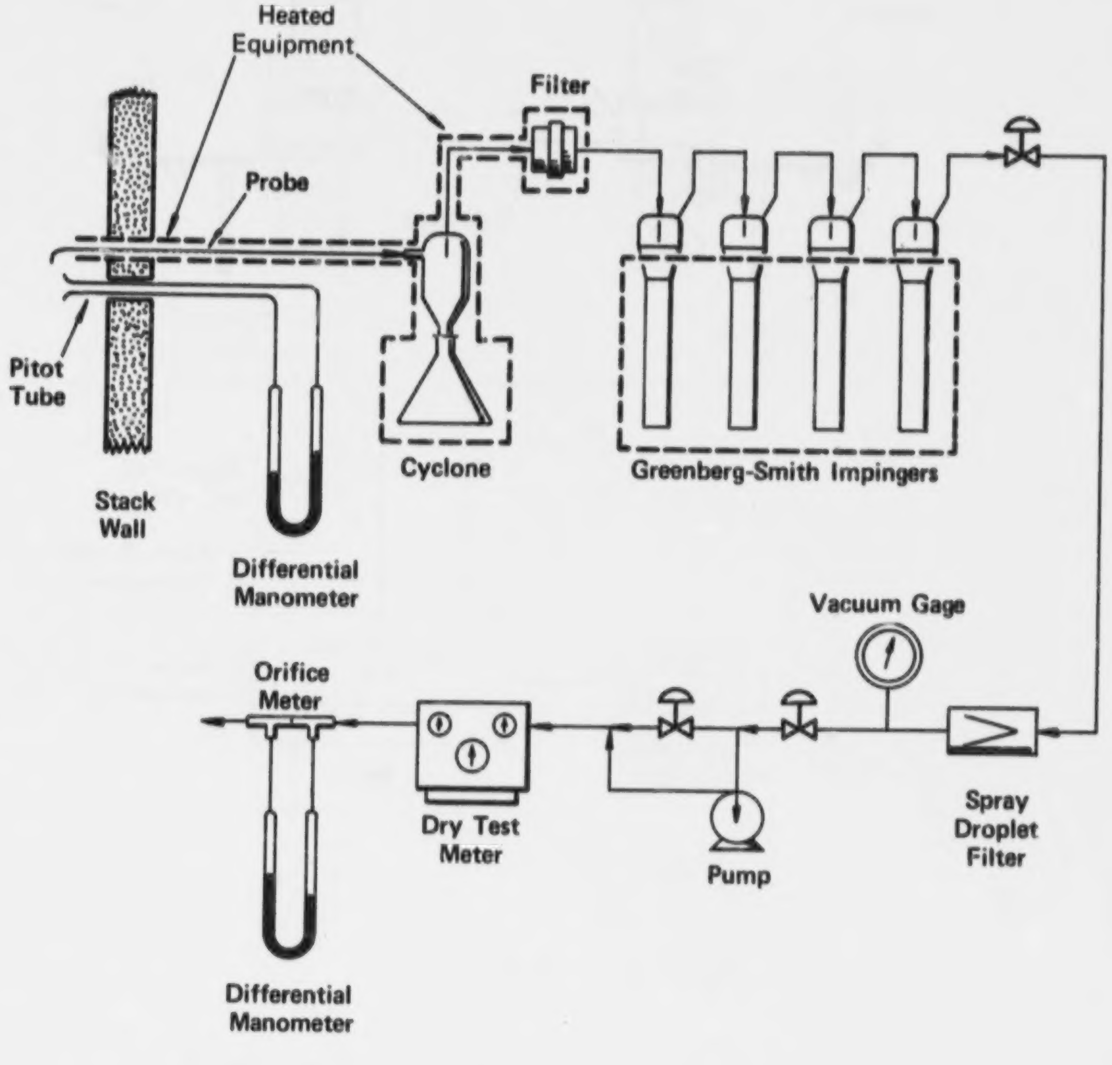

Heated Equipment
Filter
Probe
Pitot Tube
Stack Wall
Differential Manometer
Cyclone
Greenberg-Smith Impingers
Vacuum Gage
Orifice Meter
Dry Test Meter
Pump
Spray Droplet Filter
Differential Manometer

Figure 10-1. PARTICULATE MATTER SAMPLING TRAIN

158h

S-14121
67784

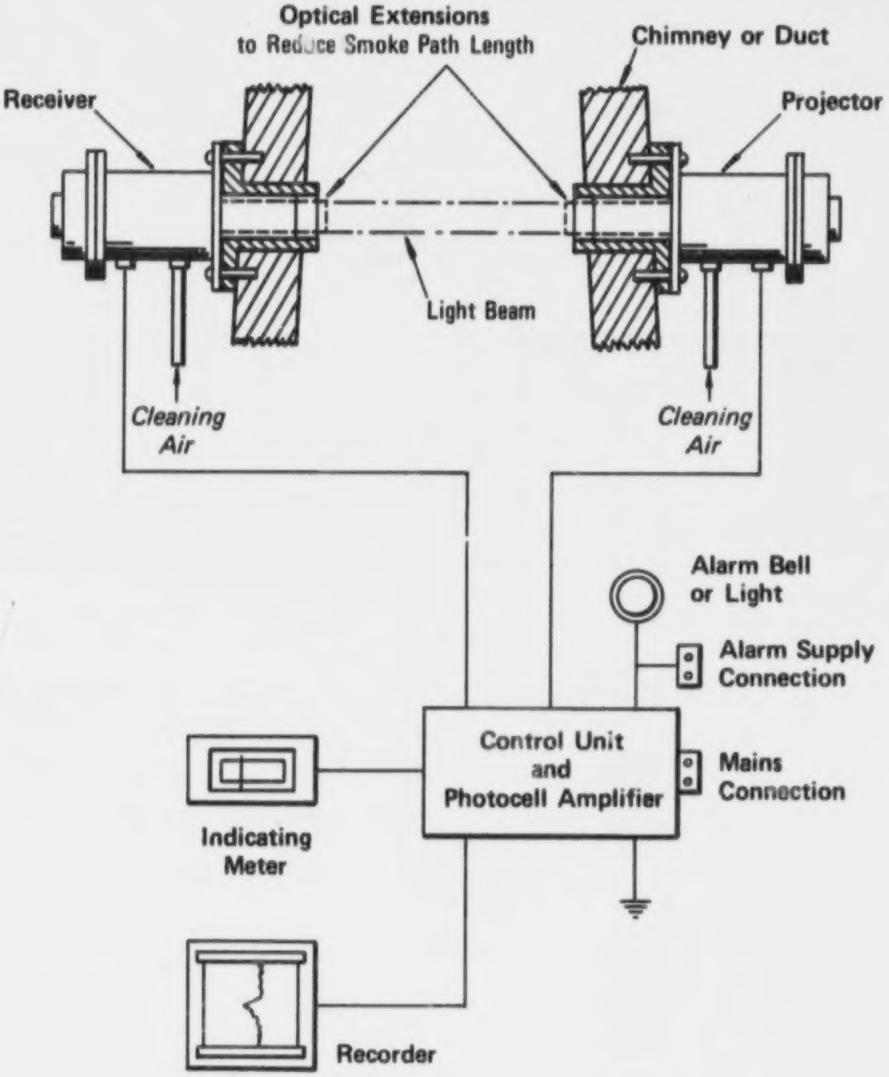

Optical Extensions
to Reduce Smoke Path Length
Chimney or Duct
Receiver
Projector
Light Beam
Cleaning
Air
Cleaning
Air
Alarm Bell
or Light
Alarm Supply
Connection
Control Unit
and
Photocell Amplifier
Mains
Connection
Indicating
Meter
Recorder

Figure 10-2. IN-STACK SMOKE PHOTOMETER

determination and measurement of odorants. No consideration will be given to the determination of odor quality or objectionability since this in itself is a very complex subject, and usually the goal in industrial odor control is the removal of the odor whether it is judged pleasant or unpleasant.

10.4.9.1 Measurement of Odor Intensity Using Dilution

There are two main concerns in the measurement of odor emissions: how serious is a given odor problem, and how effective have control measures been. To date the relative strength of an odor is generally determined by dilutions. The greater number of dilutions with odor-free air to bring an odorous gas sample to the threshold, or barely detectable, level, the stronger the odor. The strength of the odor is described in "odor units" and an odor unit is defined as the amount of odor necessary to contaminate 1 cubic foot of odor-free air to the threshold level. The number of odor units of a given sample represent the dilution with odor-free air necessary to bring the sample to the threshold level.

10.4.9.2 Sample Collection and Dilution

Samples of odorous gases are collected from the source, usually through a glass probe into a glass, plastic, or metal container. Where condensation is a problem the samples are diluted with dry, odor-free air during collection.

Once the sample is collected it is diluted to threshold level as observed by an odor panel. Many authors have described the equipment and techniques used in these operations.[10-28-40]

10.4.9.3 The Odor Panel-Determining the Odor Threshold

The number of people required on an odor panel has varied from 2 to 8. The training and use of an odor panel is at best cumbersome and expensive and there is always a tendency to use as few people as possible. The Air and Industrial Hygiene Laboratory of the State of California Department of Public Health describes an apparatus which may allow good measurements with only 2 observers.[10-34] According to ASTM Standard Dp1391-57[10-28] results that are reproducible within ±50% may be achieved with a single observer.

While these techniques are at best semi-quantitative, and more study is necessary to determine losses during dilutions, they have proved of value in judging the effects of various odor control measures (principally afterburners)[10-30,33]. Benforado[10-30] estimates that an odor strength of less than 150 odor units per scf in stack gases will not cause community odor problems. It also appears that an emission of up to about 1,000,000 odor units per minute from a plant will not cause an odor nuisance.

10.4.9.4 Possibilities for Instrumental Analysis

While the above dilution to threshold techniques can give meaningful measurements, they are not amenable to automatic monitoring. Techniques that could identify which of the pollutants are key odorants might supply information that would be useful in the design of monitors that would indicate effectiveness of odor control. One approach to this problem is to combine gas chromatography of the pollutants with sensory evaluation of the emerging resolved materials. In the study of diesel fuel exhausts, Dravnieks10-42) has made use of a GC equipped with a sniffing port at the outlet to identify key odorants. This technique is only in the development stage and is not ready for routine application, but it does offer promise for the future.

10.5 On-Site Instrumental Analysis

The present trend in analysis of gas streams is toward instrumental methods. Instruments are presently available or can be adapted for analysis of most of the components of interest. Table 10-7 summarizes the types of commercial equipment currently available.

10.6 Temperature Measurement

Measurement of temperature can be divided into four basic techniques:

1) Thermoelectric thermometry (thermocouple technics)

2) Resistance thermometry

3) Radiation and Optical Pyrometry

4) Other technics used in flame research

The first two are primarily used to study an afterburner and the last two will only briefly be described. A good basic reference on temperature and pressure measurement is Fundamentals of Temperature, Pressure and Flow Measurement by Robert P. Benedict.10-45)

10.6.1 Thermoelectric Thermometry

Thermocouples rely on the principle that junctions of two dissimilar metals generate an electrical potential which is a function of temperature (Seebeck Effect). When a closed circuit is formed with one junction at a reference temperature (usually ice water at room temperature) and the other at the temperature of interest the resulting potential difference can be measured and used to indicate temperature.10-45) Some of the most commonly used thermocouple materials are shown in Table 10-8.

Table 10-7. SUMMARY OF ANALYZERS AND MONITORS

Purpose	Manufacturer	Model	Price[a]	Address
SO_2	*Consolidated Electrodynamics	Titrilog		Pasadena, California
	*Beckman Instruments	Model 906		Fullerton, California 92634
	*Barton	286		Monterey Park, California 91754
	*Philips Electronic Instruments	PW9700	$2900	Mount Vernon, New York 10550
	*Atlas Electric Devices	1200		Chicago, Illinois 60613
	*Atlas Electric Devices	1500		Chicago, Illinois 60613
	*Leeds and Northrup	Aeroscan		Philadelphia, Pennsylvania
	*Calibrated Instruments	Wosthoff U3S		New York, New York 10023
	*Instruments Development Co.	IDC 902-1		Reston, Virginia
	*Scientific Instruments	SI-67		Hampstead, New York
	*Technicon Corporation	Auto-AnalyzerIV	$5430	Tarreytown, New York 10025
	Dynasciences Corporation	SS-130		Chatsworth, California 91311
	Dynasciences Corporation	SS-330	$2000	Chatsworth, California 91311
	EnviroMetrics	S-64	$1900	Marina Del Rey, California 90291
	Theta Sensors	LS-800-AS	$1150	Orange, California
NO_x	*Beckman Instruments	IR-315A Acralyzer NDUV-25S		Fullerton, California 92634
	*Technicon Corporation	Auto-AnalyzerIV	$5430	Tarrytown, New York 10025
	Mine Safety Appliance	LIRA-200	$2500	Pittsburgh, Pennsylvania 15208
	Dynasciences Corporation	NX-110, NX-130	$2000	Chatsworth, California 91311
	EnviroMetrics	N-76	$1900	Marina Del Rey, California 90291
	Theta Sensors	LS-800-ANX	$1150	Orange, California
	Thermo-Electron Corporation		$6500	Waltham, Massachusetts
	Bendix Environmental Science Division			Baltimore, Maryland
	REM. Incorporated	642		Santa Monica, California

a) Approximate, 1970.

Table 10-7. SUMMARY OF ANALYZERS AND MONITORS (Contd)

Purpose	Manufacturer	Model	Price	Address
Oxygenates CO	Mine Safety Appliances Co.	Lira Model 200	$2500	Pittsburgh, Pennsylvania 15208
	Bendix	Sensor Model 902	$ 575	Fall River, Massachusetts 02722
CO_2	Beckman Instrument Co.	IR-215	$2400	Fullerton, California 92634
	Mine Safety Appliances Co.	Lira Model 200	$2500	Pittsburgh, Pennsylvania 15208
O_2	Beckman Instrument Co.	CZ	$1000	Fullerton, California 92634
	Beckman Instrument Co.	715	$ 470	Fullerton, California 92634
	Teledyne	320P	$ 835	San Gabriel, California 91776
Hydrocarbons, Methane, and CO	Beckman Instrument Co.	6800	$7000	Fullerton, California 92634
	Bendix Corp.	Environmental Chromatograph		Ronceverte, W. Virginia 24970
	Mine Safety Appliances Co.			Pittsburgh, Pennsylvania 15208
	Tracor, Inc.	350F		Austin, Texas 78721
	Union Carbide	3020		White Plains, New York
	Monsanto Enviro-Chem			Dayton, Ohio
	Byron Instrument Co.			Raleigh, North Carolina
Particulates	Bendix Corp. Environmental Science Division	Tape sampler-Recorder 23605	$1212	Baltimore, Maryland 21204
	Shell Development Co.	Opacity Meter	$ 995	Oakland, California 94623
Calibration Devices	Analytical Instrument Development, Inc.	305	$ 725	West Chester, Pennsylvania 19380
	Kin-Tek Laboratories	670	$ 865	Texas City, Texas 77590
	Tracor Analytical Instruments		$2000	Austin, Texas 78721

* These instruments are primarily for ambient air concentrations

Table 10-8

Thermocouple Material	Temp. Range
Iron-Constantan	-300°F - 1390°F
Chromel-Alumel	0°F - 2490°F
Platinum-Platinum/Rhodium	32°F - 2990°F
*Tungsten 5% Rhenium-Tungsten 26% Rhenium	To 4000°F

* Requires inert gas atmosphere

All these thermocouples are available commercially in a variety of sizes, from microprobes to industrial sizes and in many sheath materials as shown in Table 10-9.

Table 10-9

Sheath Material	Notes	Melting Temperature
Stainless Steel 446	Often used in preference to Chromium-Nickel alloy in sulfur bearing atmospheres	2600 to 2750°F
Inconel* X	Nickel-Chromium alloy, age hardenable	2540 to 2600°F
Inconel* 702	Addition of Aluminum increase oxidation resistance	2400°F to 2555°F
Aluminum (type 6061)		1080 to 1200°F
Copper		1981°F
Hastelloy† C	Alloy, principally Nickel, Chromium and Molybdenum	2310 to 2450°F
Hastelloy† X	Alloy, principally Nickel, Chromium Iron and Molybdenum	2350°F
Haynes† 25	Cobalt case Alloy	2425 to 2570°F
Monel*	Nickel-Copper Alloy	2370 to 2460°F
Nickel		2615 to 2635°F
Platinum		3217°F
Platinum-10% Rhodium	Addition of Rhodium increases service temperatures, strength & corrosion resistance	3370°F approx.
Platinum-20% Rhodium		3450°F approx.
Tantalum	Useful to 4000°F in inert gas environments or in a vacuum	5425°F

* T.M. International Nickel Co., Inc. † Union Carbide Corp.

It should be remembered that a thermocouple inserted into a moving gas stream does not measure the static temperature of the gas nor the total or stagnation temperature but somewhere in between. However for most afterburner systems where the velocities are low (less than 100 ft/sec) the difference between the total and static will be less than 1°F. Of far greater importance is the effect of radiation primarily and secondarily stem conduction. Basically the temperature that the probe assumes will be due to a heat balance between convection and radiation to the probe and conduction and radiation away from the probe. The influence and importance of each of these will depend on the following factors:

a) Probe size and shape

b) Gas velocity

c) Probe thermal conductivity

d) Emissivity of the probe and surrounding

e) Fluid properties including gas emissivity and absorption

The trends of these various effects are presented in the literature as for example, in Reference 10-45 and 10-46. Typically for an unshielded thermocouple in a furnace with a gas temperature of around 2000°F and cold walls at 80°F the temperature of the probe could be as much as 40°F cooler than the gas. If the probe "sees" the flame the probe temperature could be as much as 200°F higher than the gas temperature. Thus either the probe should be designed to significantly reduce the effect of radiation or the probe should be corrected for radiation. It should be remembered too that any probe either for velocity or temperature measurement is going to disturb the system. Thus probes should be made as small as is economically possible. A properly designed probe can yield temperature measurements as accurate as ±1°F

A great variety of probe types have been developed of which several are shown in Figure 10-3.

a) Half shielded - this probe of the three is least sensitive to radiation, Mach and Reynolds Number effects.

b) Pencil type - most sensitive to radiation, Mach and Reynolds Number effects.

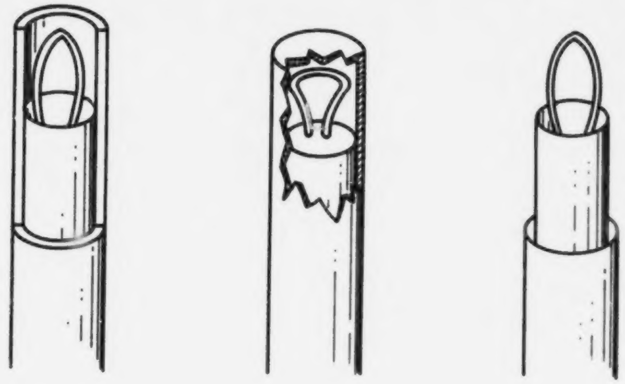
c) Bare Wire - intermediate between the above two.

These probes can also be built in the form of rakes[a)] for temperature traverses.

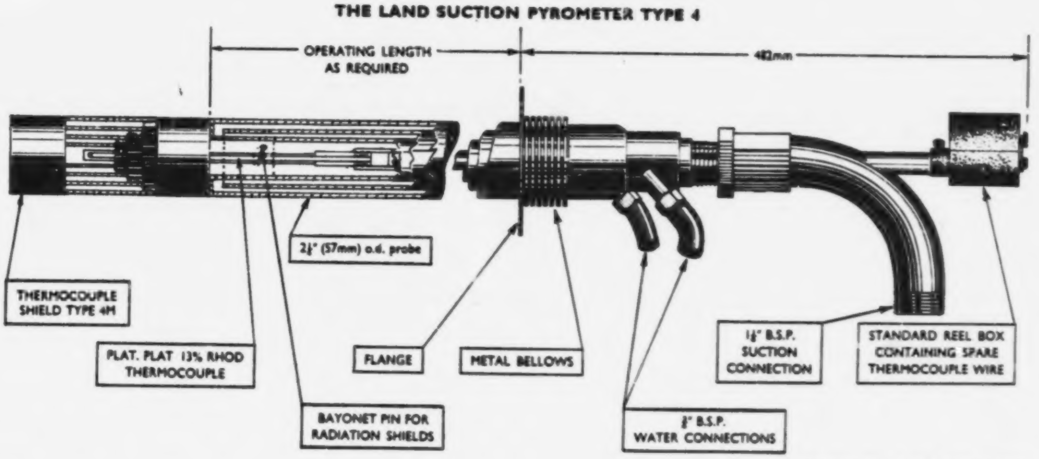
In addition some special probes have been developed for use in highly radiative regions. These are known as suction pyrometers and pneumatic pyrometers.

[a)] A rake is assembly of probes spaced at regular intervals.

Figure 10-3 follows

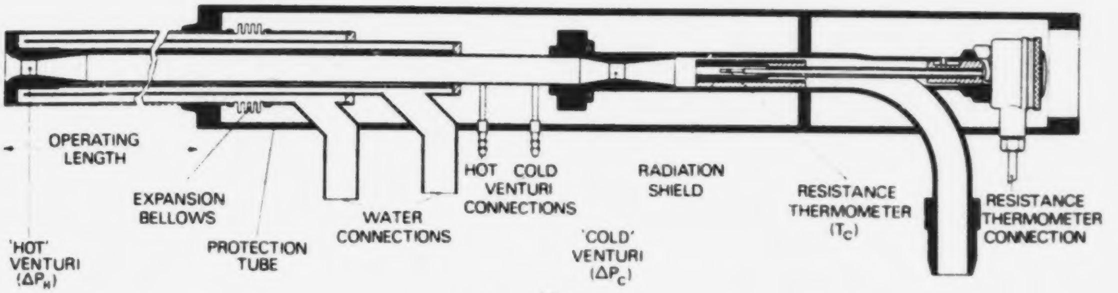


A. Half-Shielded B. Pencil C. Bare


THE LAND SUCTION PYROMETER TYPE 4
OPERATING LENGTH AS REQUIRED
482mm
THERMOCOUPLE SHIELD TYPE 4M
PLAT. PLAT 13% RHOD THERMOCOUPLE
2¼" (57mm) o.d. probe
BAYONET PIN FOR RADIATION SHIELDS
FLANGE
METAL BELLOWS
¾" B.S.P. WATER CONNECTIONS
1½" B.S.P. SUCTION CONNECTION
STANDARD REEL BOX CONTAINING SPARE THERMOCOUPLE WIRE

COURTESY: *Land Instruments*

D. Suction Pyrometer


OPERATING LENGTH
EXPANSION BELLOWS
'HOT' VENTURI (ΔP_H)
PROTECTION TUBE
WATER CONNECTIONS
HOT COLD VENTURI CONNECTIONS
RADIATION SHIELD
'COLD' VENTURI (ΔP_C)
RESISTANCE THERMOMETER (T_C)
RESISTANCE THERMOMETER CONNECTION

$$T_H = K \frac{\Delta P_H}{\Delta P_C} T_C$$

COURTESY: *Land Instruments*

E. Section of Land Venturi Pneumatic Pyrometer, Type VPP15

Figure 10-3. THERMOELECTRIC PROBES

1642

A suction pyrometer (see Fig. 10-3d) is a device that completely shields the thermocouple junction from radiation and draws the hot gases over the junction at velocities in excess of 150 ft/sec. Accuracies of 1-1/2% to 2% are typical. Some of these instruments are water cooled and typically operate to 2900°F. They are available commercially.

To make measurements in the flame region to 4000°F pneumatic pyrometers have been developed. Pneumatic probes are generally of two types, venturi pneumatic pyrometers (see Fig. 10-3d) and sonic orifice pneumatic probes. Both of these probes rely on the thermodynamic relationships of gas flow across a series of sonic orifices or venturies. In the venturi probe (see Fig. 10-3e) aspirated hot gas density is measured by a front venturi restriction; after being cooled by passage down the water cooled probe, the density is measured by a second venturi and the cool gas temperature is measured by a thermocouple or resistance thermometer. Generally these devices are used where high temperature, dirty gas conditions or faster response requirements would prohibit the use of suction pyrometers. However, care must be exercised if a double sonic orifice is used in a dirty environment since corrosion or deposition can take place. It has also been reported that it is difficult to interpret the results when there are rapid fluctuations in pressure, temperature and velocity.[10-47] Again these probes are available commercially and can be made to perform traverses either radially or axially.

In summation the following advantages and disadvantages may be listed for thermoelectric thermometry:

Advantages	Disadvantages
a) High precision characteristics of electrical measurement	a) Radiation error correction needed (Can use null technique to eliminate radiation)
b) Extremely small thermocouple can be made	b) Spatial resolution limited to wire size
c) Can be made of materials to withstand high temperatures.	c) Probe can generate disturbances due to vibration or catalysis (The latter may be eliminated by coating the T/C with silica[10-46])

10.6.2 Resistance Thermometry

Resistance thermometry relies on the principle that a wire's resistance changes in a known manner with temperature. Basically there are three characteristics of these devices: 1) Simplicity of the circuit (no reference junction is needed) 2) Sensitivity of measurement and 3) Stability of the sensors. There are two basic types of sensors, resistance temperature detectors (RTD's) and thermistors. RTD's are solid wire probes, platinum, nickel or copper typically and have a positive coefficient of resistivity. They have a temperature limit of 1000°F. Thermistors which have a negative coefficient of resistivity are electrical circuit elements formed from solid semi-conducting materials. They are quite accurate,

±0.1°F but are limited to a temperature range of 32°F - 600°F. These temperature ranges are rather low for direct application in an afterburner but the elements are used as secondary elements in the pneumatic pyrometers (Section 10.6.1) for greater accuracy.

10.6.3 Radiation Pyrometers

There are two classes of radiation pyrometers, total radiation and optical pyrometers. Total radiation pyrometers measure the total radiation emitted by a hot body. Optical pyrometers compare the brightness of the hot body in a narrow frequency or color band with the brightness of a standarized radiation source. Typically these could be used to measure the refractory surface temperatures. Many instruments in most any price range and accurracy are available commercially. Table 10-10 shows a typical relation of color to temperature of iron or steel.

Table 10-10. RELATION OF COLOR TO TEMPERATURE OF IRON OR STEEL

	Deg F
Dark blood red, black red	990
Dark red, blood red, low red	1050
Dark cherry red	1175
Medium cherry red	1250
Cherry, full red	1375
Light cherry, light red	1550
Orange, free scaling heat	1650
Light Orange	1725
Yellow	1825
Light yellow	1975
White	2200

10.6.4 Other Techniques

Many other sophisticated techniques have been used for detailed flame studies such as the sodium line reversal method, spectral line intensity methods, and temperatures derived from measurement of gas density gradients. The reader is referred to the book by Fristrom and Westenberg (Reference 10-46) for the details of these techniques.

10.7 Flow Pattern Determination

Two approaches may be taken in attempting to determine flow patterns in an afterburner. One can study the full scale system either hot or cold or one can study a geometrically similar model.

10.7.1 Use of Models

Typically, a plastic model would be employed with water or air as the fluid in order to simulate afterburner flow patterns. Care must be exercised in modeling to insure that the important parameters are held constant. That is, the experimenter must decide whether velocity ratio, Reynolds number, density ratio and/or some other dimensionless groups are the most important characterizing parameters. A good discussion of this is given in Reference 10-49. By using a model, considerable time and expense can be saved and a good qualitative feel for the flow patterns can quickly be ascertained through visual techniques. Quantitative methods for determining flow patterns are also available and can be used with either cold models or full scale systems.

Water models have been very popular for simulating furnaces and combustors because of the ease of pumping water and the many qualitative and quantitative visual techniques which can be used to determine flow patterns (Section 10.7.3). It should be remembered though that with a water model, an incompressible fluid is being used to simulate compressible fluid flow. Mach number effects and expansion due to heating cannot be adequately simulated. However, in afterburners velocities are generally quite low and compressibility effects will not be too important. Also, good correlation has been obtained between water model studies and observations in operating gas turbine combustors$^{10-50)}$, industrial furnaces and domestic heating equipment.

10.7.2 Quantitative Methods

Velocity measurements are usually made using pitot tubes or hot wire anemometry. The latter method is only suitable for cold studies. For measurements in highly turbulent flows, a straight pitot tube is difficult to use successfully since the orientation must be within a few degrees of the flow direction. Thus, a variety of multihole probes have been developed to measure velocity and flow direction in either two or three dimensional systems. Examples of 2-D probes are claw probes, three hole yaw probes, cobra probes and wedge probes (See Fig. 10-4a-d). The procedure is simply to balance the two outer pressure taps to orient the probe and then make the velocity measurement through the center tap. Suitable methods of measuring the angle to some reference line and traversing micrometer must be employed in order to map the flow chamber. Three directional probes are the Four or Five Hole Yaw Probe, Spherical probe, five hole directional probe and 3-D cobra (See Fig. 10-5a-c). All of these are described in detail in Reference 10-51. Many of these probes can be cooled for use in hot systems and are available commercially.

Hot wire anemometry can also be used for velocity measurement, flow direction and turbulence intensity and is particularly good in measuring unsteady flows. It is basically an indicator of rate of heat transfer from a hot wire to the flowing system. The reader is referred to Chapter VI of Aerodynamic Measurements, Reference 10-51 for a very complete discussion of the fundamentals of hot wire anemometry. Hot wire anemometry has primarily been used to study boundary layers and turbulence. See also Reference 10-52. A few tracer materials have been used successfully such as helium or hot air injection where the concentration levels can be mapped with a thermal conductivity cell. These concentration levels then give an indication of the amount of mixedness and indicate flow patterns. These techniques are somewhat limited if the system is in an unsteady state as the thermal conductivity cells will indicate time averaged values. In addition if the system is highly turbulent with a great deal of back mixing it may be difficult to obtain a good concentration map. Radiosotopes have been used in a similar manner.

Many of the visual techniques described in Section 10.7.3 can be used in conjunction with photography to yield quantitative information.

10.7.3 Visual Methods

Visual methods in conjunction with air or water models are probably the least expensive and fastest way to obtain a great deal of qualitative information. (As is discussed below, it is also possible to get some quantitative information.) Design changes can be easily made on a model and visual observations made to assess its effect on flow patterns. Visual methods may also be used in the full scale apparatus, either cold or at operating temperatures. Visual techniques may be divided into two categories, static methods and kinetic methods.

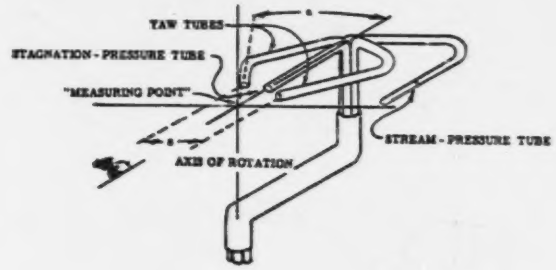
10.7.3.1 Static Methods

Static techniques are most often used for boundary layer flow visualization and may involve a surface coating (oil paint, dye, or fluorescent oil film[10-53]) or tufts attached to the wall for directional measurements (regions of reverse flow may be detected)[10-54]. This gives a general picture of flows in the region near the surface under observation but gives little if any information about disturbance in the main flow. For this reason, static methods would not be used for afterburner flow pattern studies.

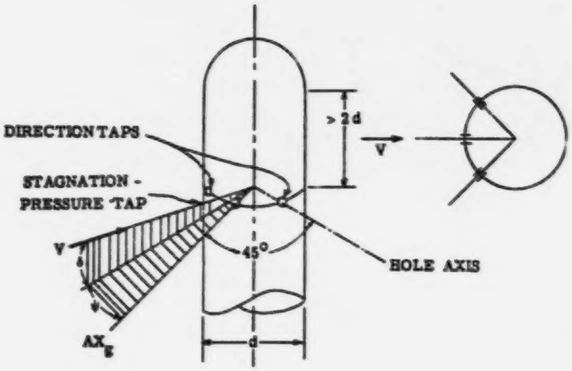
10.7.3.2 Kinetic Methods

The most direct method for flow visualization is with suspended microscopic dust particles. Flow is made visible in a cross section by a high-intensity, narrow plane beam of light cutting a slice across the flow path and can be viewed or photographed at right angle to the plane beam. Table 10-11 lists some of the common tracers for air or water systems. For good photographic records the particles should be as large as possible and for hydrodynamic reasons as small as possible. In addition the tracers should be as near as to a neutral density as possible so as not to be influenced by buoyancy, inertia or gravitational forces. Great care must be exercised in preparing the background, selecting a strong light source, and if quantitative information is desired a suitable technique for pulsing either the tracer or the illumination for a photographic record. These aspects are discussed in great detail in References 10-46, 54, 55, 57. Probably the easiest to use of all these techniques (limited to water models) is the hydrogen bubble technique or air bubbles. It has the advantage particularly in a closed system that particles do not obstruct any pitot tubes that happen to be in the system.

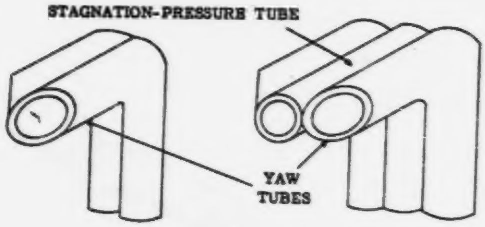
Figures 10-4 and 10-5 follow
Table 10-11 follows


YAW TUBES
STAGNATION - PRESSURE TUBE
"MEASURING POINT"
STREAM - PRESSURE TUBE
AXIS OF ROTATION

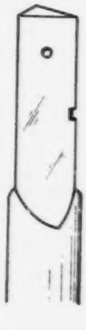
A. Claw Probe


DIRECTION TAPS
STAGNATION - PRESSURE TAP
> 2d
V
V
45°
HOLE AXIS
AX_g
d

B. 3-Hole Yaw Probe

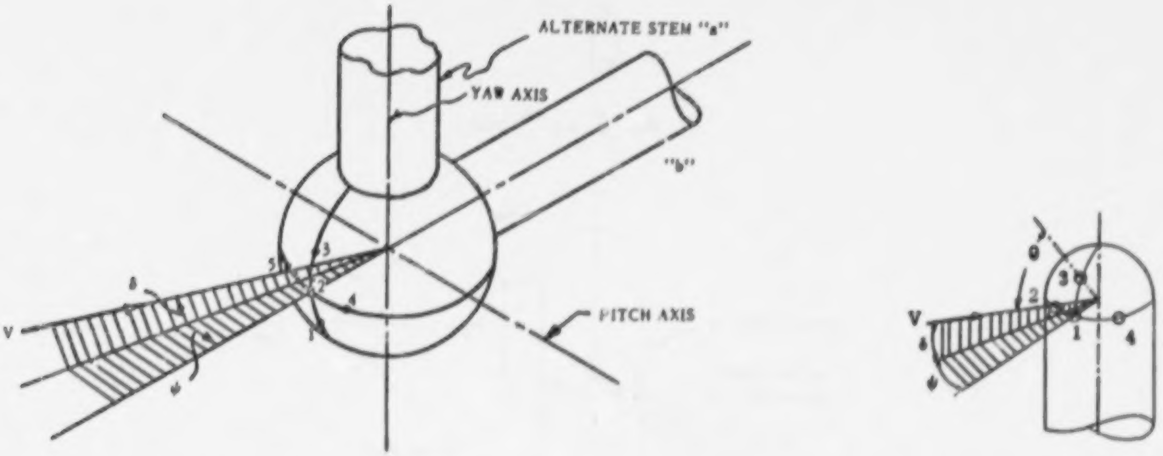

STAGNATION-PRESSURE TUBE
YAW TUBES

C. Cobra Probes



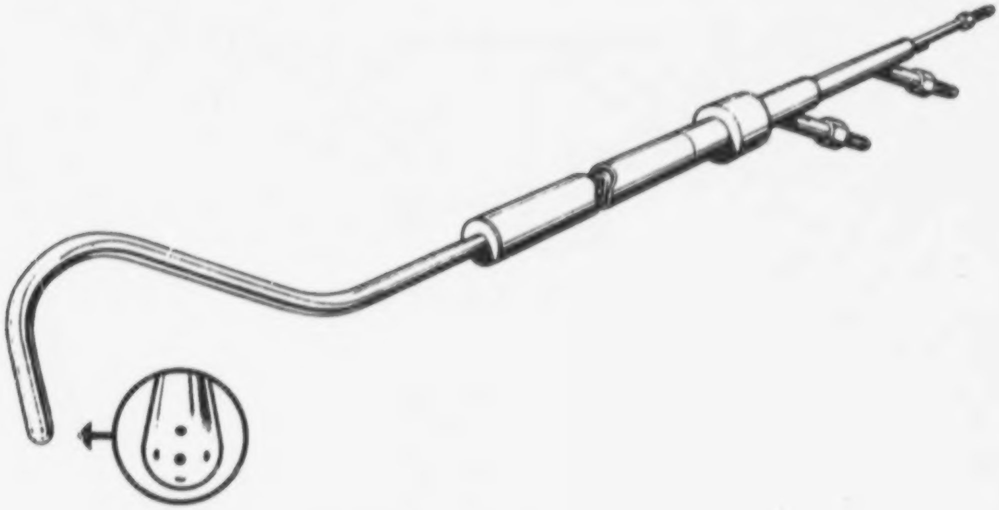
D. Wedge Probe

Figure 10-4. TWO DIMENSIONAL PROBES


ALTERNATE STEM "a"
YAW AXIS
"b"
PITCH AXIS
V
δ
ψ
1
2
3
4
5
θ

A. Spherical Yaw Probe

B. 4-Hole Yaw Probe



C. 5-Hole Directional Probe

Figure 10-5. THREE DIMENSIONAL PROBES

Table 10-11

References	Material	Dispersed In Air = A Water = W	Melting Point °F	Boiling Point °F	Spec. Gr. @ 60°F	Particle Size	Suitable In Hot = H Cold = C	Comments
10-55,54,46,49	Al	A, W	1220	3735	2.7		C	
10-56	NH_4Cl	A	Sublines	968	1.53		C	
10-56	$TiCl_4$	A	>3272		5.47		H,C	Forms TiO_2 in flames. Sublimes and is reduced by flame gases.
10-55,54,49	Polystyrene	W	148-200		.98 - 1.1	4-14 mills	C	
10-49	Balsa Dust	A			.1 - .2		C	
10-56,49	Smoke	A					C	Typically boiling oil vapors . Not particularly good in turbulent flow regimes.
10-58	Metaldehyde-Flocks	A				2 mm		Good light scatter.
10-55,54	Air Bubbles	W			.0012		C	Difference in densities most serious, drawback, electrochemical technique, Sec Reference 55 & 57. Used mostly for 2-D studies. Easy to get pulsed traces.
10-55,57	O_2	W			.0013		C	
10-55,57,54	H_2	W					C	
10-59,55,57,54	Dye Streaks	W					C	Primary laminar flow; difficult to use if a closed system. See reference for actual dye materials.
10-56	Al_2O_3	A	3632	4010	4.0		H	
10-56	Pyrex	A	752		2.24		H	Very reflective
10-56	MgO	A	5072	6512	3.65	1-5 u	H	Generally considered best.
10-56	Quartz	A	3092	4046				
10-56	TiO_2	A	3866	Decomposes	3.84-4.17		H	
10-56	ZrO_2	A	5432	7772	5.73		H	

Chapter 11. COSTS

11.1 Scope

This chapter covers the purchase, installed, and operating costs of thermal and catalytic afterburners with and without heat exchangers. The information is based on extensive discussions with manufacturers and users, and also on user replies to our questionnaires. The data have been presented in graphical form so that they may be used easily.

Fuel requirements are discussed in detail including the effects of heat exchange and of hydrocarbons in the waste gas stream. Comparisons have been made between the total annual costs, including depreciation, for typical thermal and catalytic units with varying levels of heat exchange.

11.2 Thermal Afterburners

11.2.1 Purchase Costs from Manufacturers' Data

Purchase costs obtained from sixteen manufacturers of thermal afterburners are plotted as broad bands in Figure 11-1. They cover units without heat exchangers, with 35 to 45% heat recovery, and with 55 to 75% heat recovery. The range in costs for a particular type of unit may be accounted for by different methods of construction, type of materials used, and instrument package. For instance, a simple burner in an unlined stack is sometimes used for odor control. It has an absolute minimum of instrumentation, short residence time, low operating temperature, low cost, and although it may reduce odor and plume, its performance expressed as percentage reduction in hydrocarbons would not meet stringent requirements such as Rule 66. The more expensive units are designed with 2000°F refractory lining, are made of heavy gage metal with adequate expansion joints, have baffles for good mixing and an expensive instrument package. Heat exchangers are generally shell and tube design, and the cost depends on materials of construction (e.g., aluminized tubes, carbon steel shell, versus stainless steel), the number of passes and provisions for expansion. Rotary regenerative exchangers and packed bed regenerative exchangers are sometimes used for high heat recoveries; both tend to be expensive. It must be emphasized that a unit high in first cost does not guarantee that it will have superior performance. Bids from manufacturers must be first evaluated on the ability of the afterburner to do the job, (i.e., residence time, operating temperature, mixing, etc.), and then compared as to construction details and costs.

11.2.2 Instrumentation

Instrumentation requirements to meet safety codes have been discussed in Chapter 9. For a gas fired unit a minimum cost for temperature control would be $750 with an additional $1,400 for combustion safeguards to meet FIA safety requirements (total $2,150). For more accurate and reliable pneumatic control instruments, also using a UV detector rather than a flame rod for shutdown protection, the cost would be increased to $3,600. These costs can be much higher on large systems with built-in control features and alarms to meet special process requirements with complicated start-up and shutdown procedures. Oil-fired systems require additional controls which are discussed later under burners.

Instrumentation costs can run from as low as 5% of the purchase cost to as high as 30%. On small units, in particular, they can be a substantial portion of the price. Thus, it is extremely important to compare instrument packages on different bids on an equivalent specification basis.

11.2.3 Burners

The bare cost of oil or gas burners without auxiliaries or instrumentation is relatively small. Distributed raw gas burners give better mixing of the waste gas but are more expensive than single nozzle-mix burners. Dual fuel gas/oil burners are only slightly more expensive than single fuel burners. Costs, of course, depend on the particular manufacturer, design, and materials of construction. Some afterburner suppliers use standard burners from burner manufacturers, whereas others design and build their own. The following figures were quoted to us by burner suppliers and are reproduced here to indicate the approximate burner cost in relation to the overall cost of the afterburner. They should not be considered as firm prices:

Type of Burner	MM Btu/hr	Bare Cost of Burner
5-inch nozzle mix gas	3.25 - 5	$318
Distributed raw gas	3.5	$525
4-inch dual fuel gas/oil	3.4	$260

Thus, the difference in cost between oil fired, gas fired, or dual fuel fired systems is primarily due to additional instrumentation and auxiliaries for fuel atomization including air blowers, pumps and heat exchangers. Manufacturers have indicated that a unit burning Number 2 or Number 4 fuel oil would cost $2 M - $4 M more than a corresponding gas fired installation. For burning Number 6 fuel oil, assuming that steam is available for heating the oil to reduce its viscosity, an incremental cost of $3 M - $6 M could be incurred over a gas fired unit.

It must be emphasized that information on instrumentation and auxiliaries is extremely dependent on the details of the installation. Firm prices must be obtained on a strictly comparable basis in order to be able to evaluate the relative merits of proposed burner installations by different manufacturers.

Figure 11-1 follows
Table 11-1 follows

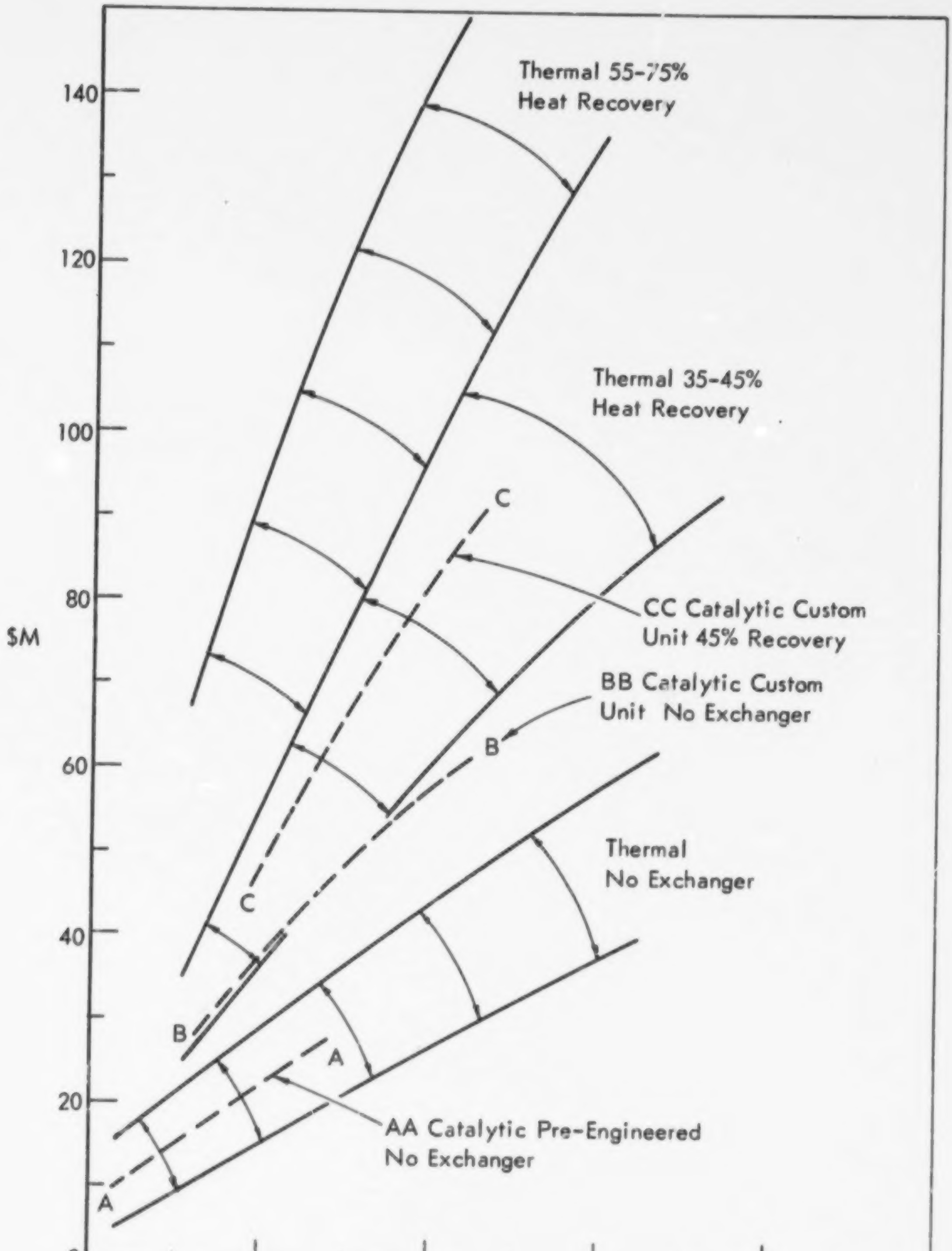


Figure 11-1. AFTERBURNER PURCHASE COSTS FROM MANUFACTURERS' DATA

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11.2.4 Purchase, Installed and Operating Cost from Users' Data

A comprehensive summary of the data obtained from users of afterburners has been tabulated in computer format and is given in Table 11-2. Items which have been included, and are of direct interest with respect to costs, are (a) capacity of the unit in thousands of standard cubic feet/minute (MSCFM), (b) purchase cost, (c) installed cost, (d) year of installation, (e) operating cost, and (f) hours of operation per day.

A computer program was written to sort the data into categories, apply corrections, and machine plot the treated data. It had the following features:

Purchase and installed costs were adjusted to a 1970 cost base by applying factors from the Chemical Engineering Fabricated Equipment Index.

Operating costs were corrected to a 24-hr/day basis. (A correction based on the number of hours worked per year would have been better, but was not available in most cases).

The data were separated into units with and without exchangers.

Six separate plots were generated covering purchase cost, installed cost, and operating cost versus capacity, for units with and without heat exchangers. They are shown in Figure 11-2 through 11-7.

Although there is a large amount of scatter in the data, this is not unexpected because of the varied conditions under which the units operate, and the wide differences in design. Most of the purchase costs fall within the broad bands of prices obtained from the manufacturers.

Installed costs for units without heat exchangers range from 1.2 - 10.0 times the purchase cost with an average of 2.2 times (obviously the ratio of 10 is exceptional; 85% of the units are covered by a ratio of 3 or less). For units with exchangers the range is from 1.2 times to 3.7 times with an average of 1.9. Thus for an average plant the installation cost is of the same order as the purchase cost. The lower installation costs are associated with smaller afterburners shipped as package units requiring little more than a concrete pad and gas supply to be furnished on site. The more expensive units are associated with large equipment, often installed on special steel supporting structures, with long service lines and interconnecting ductwork, and an expensive instrumentation package interconnected with the process.

Operating costs are largely fuel costs since maintenance is a relatively small item except in a few cases where the unit was unsuitable for the application. There have been instances where it has been necessary to do extensive repairs on an afterburner after only one year in service due to cracks in the afterburner shell and heat exchanger. This was usually due to inadequate design, insufficient allowance for expansion, and in some cases, over-temperature operation. Fortunately, such incidents represent a small number of afterburners.

Most companies do not keep separate detailed records of maintenance costs on afterburners so that the information reported to us has been meager. Eight different units without heat exchangers had an average annual maintenance cost of 1.9% of installed cost, whereas 11 units with heat exchangers had a corresponding figure of 3.5%. The higher maintenance on heat exchangers is largely due to the requirement to replace burned out tubes from time to time in some applications. It should be noted that the 3.5% factor is applied to a larger installed cost for the same throughput since units with exchangers are considerably more expensive than units without. Choice of materials for heat exchangers can make a large difference to the frequency of repair of these units. The substitution of SS tubes for aluminized tubes may be paid off in reduced maintenance cost over a relatively short period in some cases. In view of the fact that many companies reported negligible maintenance costs, it is felt that the 3.5% factor may be on the high side for a well designed system, and the correct figure may be closer to the 2% for units without exchangers. This would cover minor repairs to the refractory lining, maintenance of control instruments and cleaning of flame rods, if used.

Afterburners are often brought on stream without using special slow heat-up cycles, although some manufacturers do recommend them. Operator attention is minimal and consists of starting up the unit, noting by inspection of the flame that it is operating satisfactorily, and then relying on the temperature controller to take over. Usually the operator in overall charge of the process equipment is responsible for the afterburner, and additional operators have not been required after installing an afterburner on an existing process unit.

Fuel costs are discussed in some detail below, but for comparison purposes lines have been drawn on Figures 11-4 and 11-7 showing typical fuel costs. They are based on 6,000 hours per year operation, heating from 400°F to 1400°F, with gas at $0.50 per MSCF. The waste gas has been assumed to have no heating value, and for the case with a heat exchanger, 45% recovery has been assumed. Fuel costs lower than these curves may be explained as being due to the heating value of the contaminants in the waste gas; higher costs could be due to the local cost of fuel.

11.2.5 Fuel

Most afterburners use natural gas as a fuel since it is easy to handle and control, it burns cleanly, and is free from sulfur. It is presently low in cost because of price regulation, but unfortunately it is in short supply. It is expected that the price will be allowed to rise in the future to meet the supply-demand situation, and also to encourage exploration for new gas fields. Alternative fuels are numbers 2, 4, and 6 fuel oil with low sulfur content, which are also in short supply. Prices of both natural gas and fuel oil vary depending on the geographic location. Fuel oil prices also fluctuate seasonally reflecting demand, and the local supply situation. A typical natural gas price would be of the order of 50c/1000 SCF, equivalent to 56c/MM Btu (low heating value 900 Btu/SCF). Fuel oil could be of the order of $5.00 per barrel, equivalent to about 89.5c/MM Btu, (low heating value 133,000 Btu/gallon, 1 barrel = 42 gallons). It must be emphasized that both these costs can vary widely and local prices must be obtained when the economics of an installation are being considered.

Figures 11-2 through 11-7 follow

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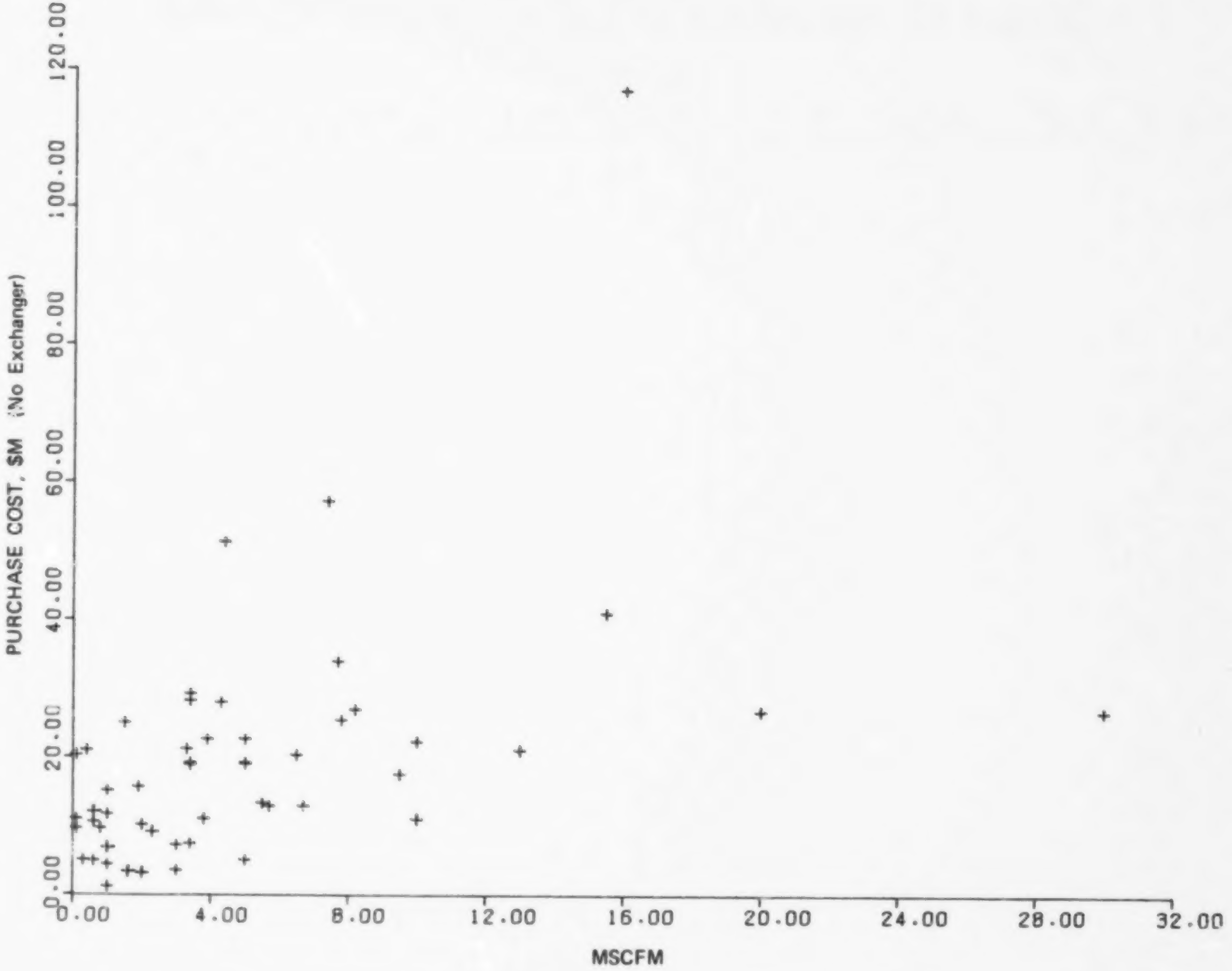

Figure 11-2. PURCHASE COST THERMAL AFTERBURNERS WITH NO EXCHANGER
(Data Taken From Users Survey)

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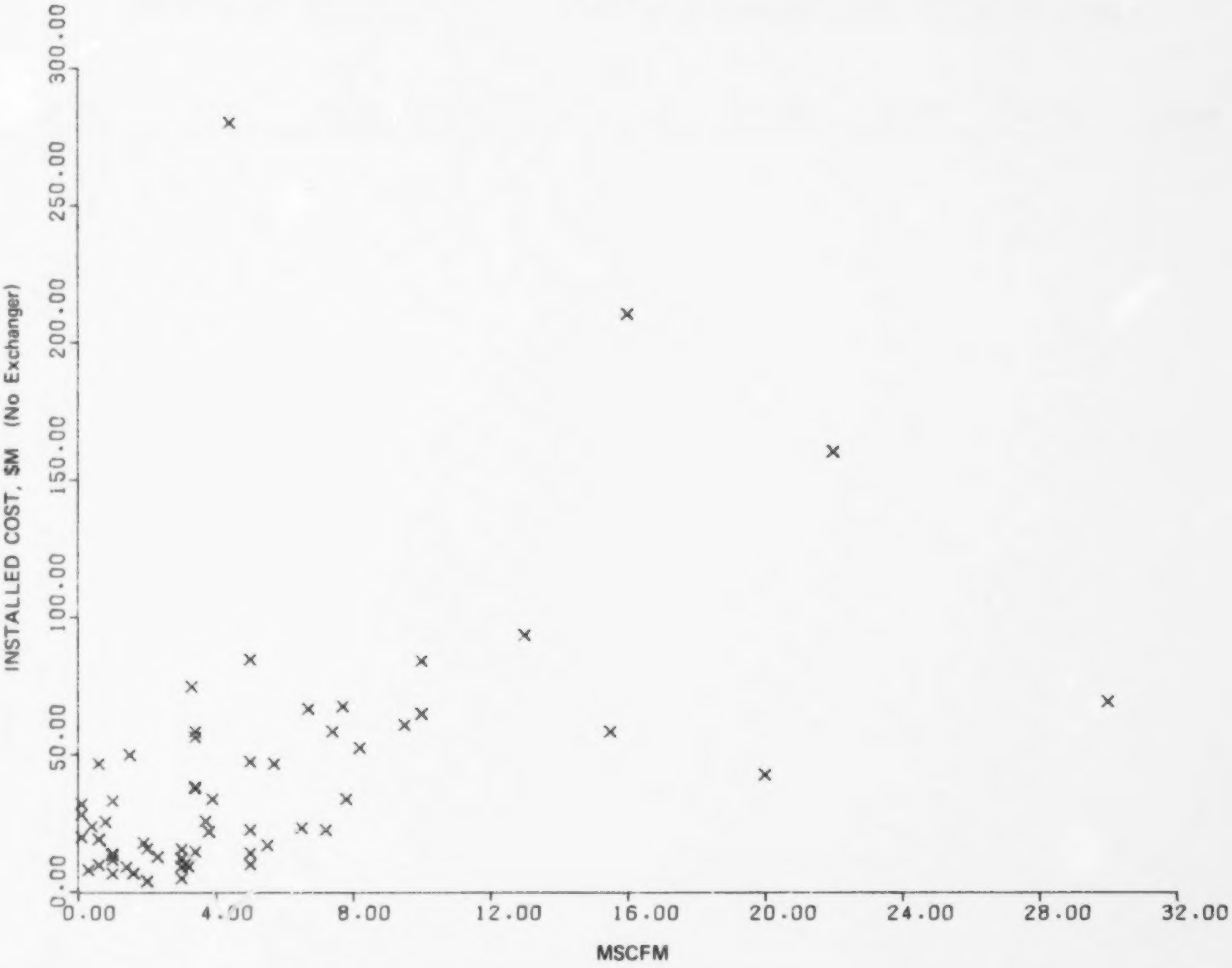

Figure 11-3. INSTALLED COST THERMAL AFTERBURNERS - NO EXCHANGER
(Data Taken From Users Survey)

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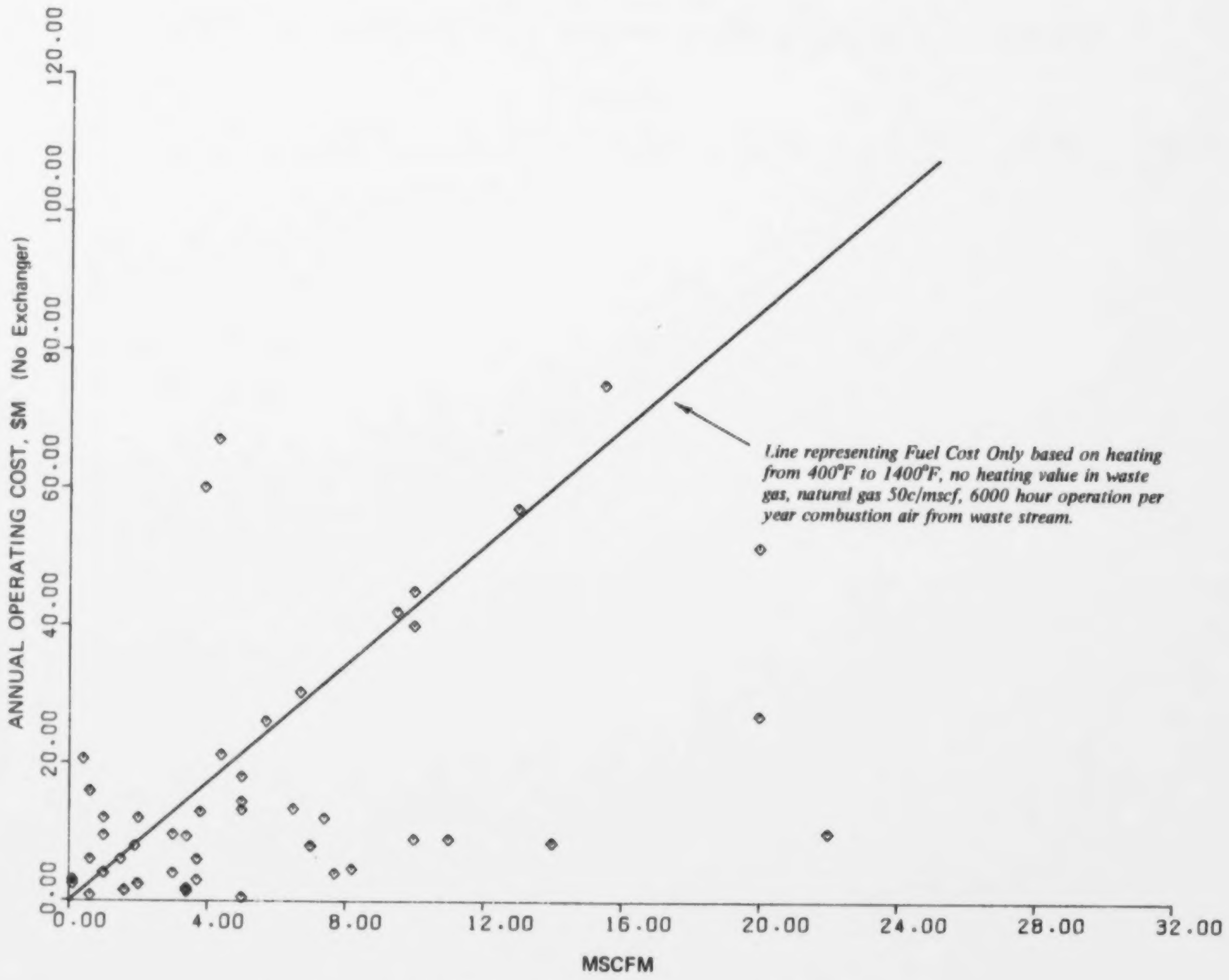

Line representing Fuel Cost Only based on heating from 400°F to 1400°F, no heating value in waste gas, natural gas 50c/mscf, 6000 hour operation per year combustion air from waste stream.

Figure 11-4. ANNUAL OPERATING COST THERMAL AFTERBURNERS - NO EXCHANGER
(Data Taken From Users Survey)

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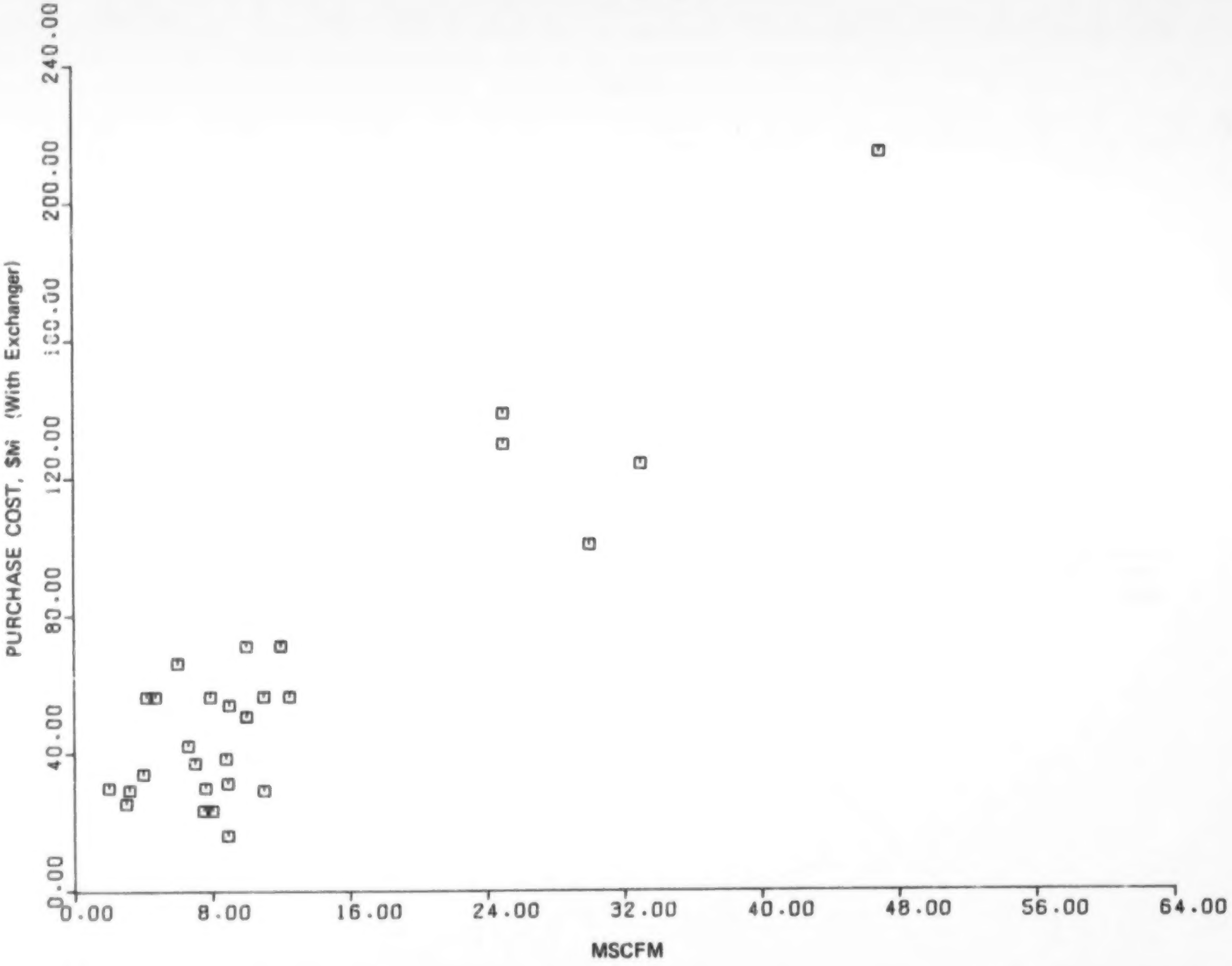

Figure 11-5. PURCHASE COST THERMAL AFTERBURNERS WITH EXCHANGER
(Data Taken From Users Survey)

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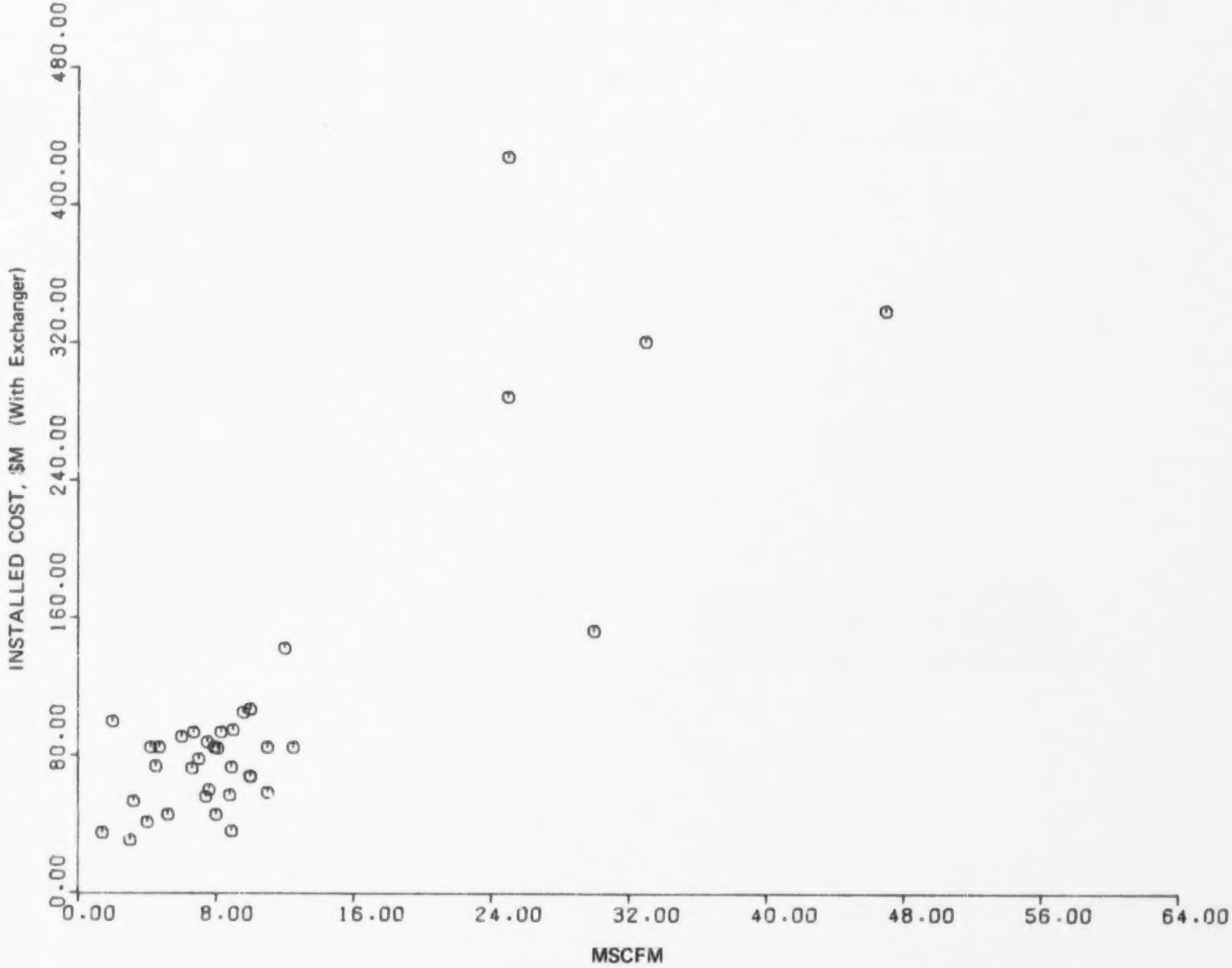

Figure 11-6. INSTALLED COST THERMAL AFTERBURNERS WITH EXCHANGER
(Data Taken From Users Survey)

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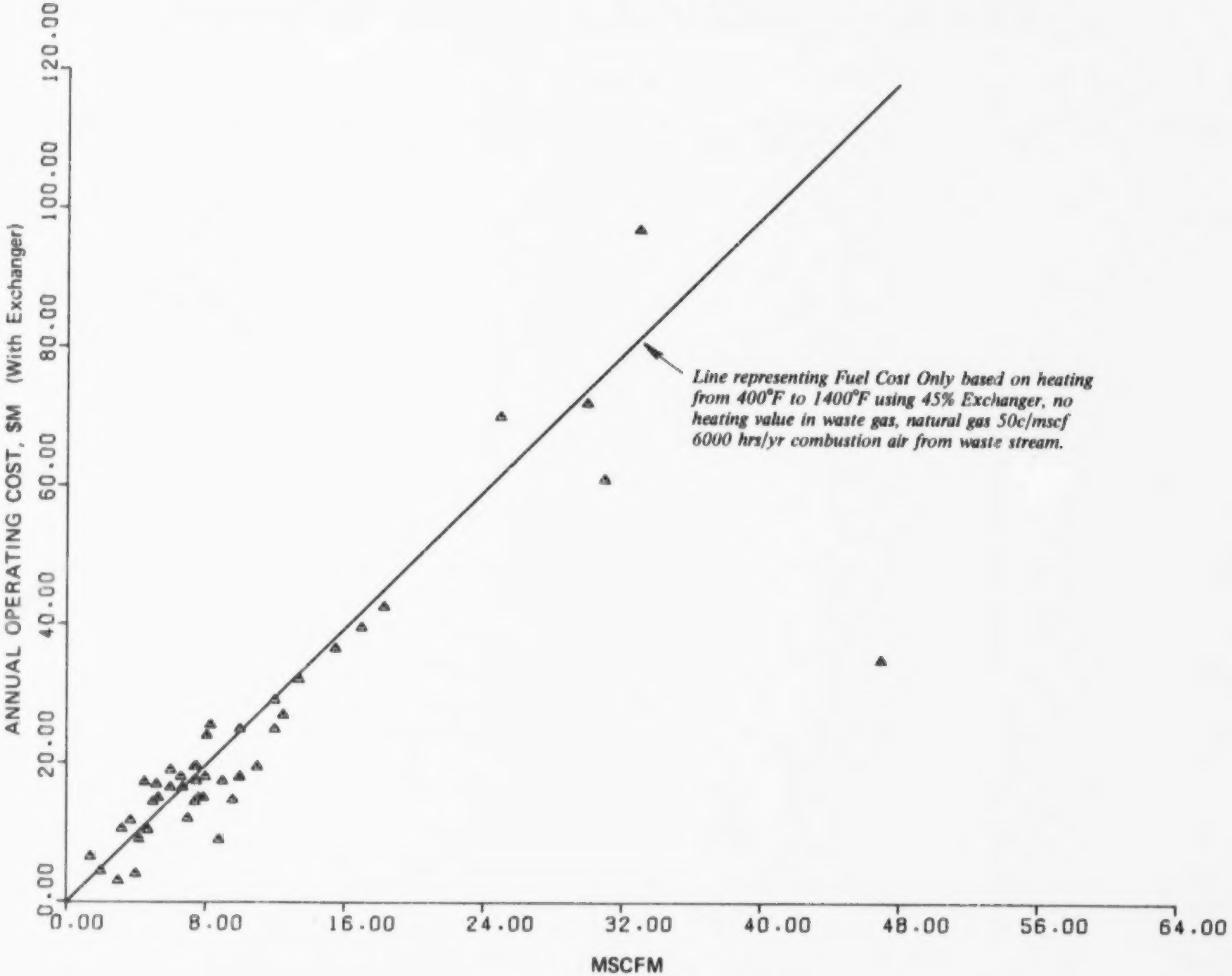

Figure 11-7. ANNUAL OPERATING COST THERMAL AFTERBURNERS WITH EXCHANGER
(Data Taken From Users Survey)

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In order to obtain a 90% reduction in hydrocarbons, thermal afterburners generally need to operate at 1400°F. Thus, if advantage is taken of the heat in the discharge from the afterburner to preheat the incoming waste gas, a considerable fuel saving can be made. Figure 11-8 shows a diagram of such a system in which the waste gas is preheated from T_A to T_B through the exchanger, the afterburner temperature is controlled at 1400°F, and the hot gas from the afterburner gives up its heat through an indirect heat exchanger to the incoming gases. A commonly used expression for the heat exchanger recovery is:

$$\text{Simplified exchanger recovery} = \frac{T_B - T_A}{1400 - T_A}$$

(A more accurate criterion of heat exchanger performance is the heat exchanger effectivness ratio discussed in Chapter 6).

Plots have been made in Figure 11-8 of the temperature T_B against exchanger recovery for incoming waste gas temperatures (T_A) of 100°F to 400°F. If temperature T_B is found in this way, the heat to be supplied by the afterburner may be calculated, and fuel requirements estimated, as is discussed later. The economic incentive for using a heat exchanger with a particular recovery may thus be obtained.

Fuel requirements are determined by the temperature rise through the afterburner, the heating value in the waste gases, the specific heat of the waste gases and whether combustion air for the fuel is brought from outside. Afterburners are often used on waste gases containing only a dilute concentration of hydrocarbons and in these cases, the oxygen in the waste stream can be used to supply all that is needed for combustion. This can amount to a 30% reduction in fuel requirements compared with bringing in the stoichiometric amount of air required for combustion from outside.

In calculating the heat requirement in the afterburner, the fact that the combustion products have quite different specific heats from the gases entering the afterburner is often neglected. Figure 11-9 shows the average specific heats of a number of gases and how they change with temperature. Items of particular interest are the values for air and carbon dioxide.

The concentration of hydrocarbons in the waste stream has a large effect on fuel requirements. For example, hexane at the 25% LEL has a heat content of about 13 Btu/SCF which would be sufficient to raise the temperature of the waste gas by about 650°F.

In order to allow for these various factors, equilibrium calculations were made using a computer program to determine the fuel requirements to reach 1400°F in the afterburner for four main cases. These included natural gas or fuel oil with either stoichiometric air from outside or combustion air supplied by the waste stream. In each case the calculations were done to cover a range of inlet temperatures to the afterburner from 100°F to 900°F with a range of hexane in the waste gas from 0 ppm to 3000 ppm. (The amount of hydrocarbon in the waste stream is commonly reported either as parts per million carbon (ppm C_1) or parts per million hexane. Since hexane contains 6 carbon atoms, ppm C_1 = ppm hexane x 6.) The results are plotted in Figures 11-10 through 11-13. (For the purpose of these calculations the natural gas was assumed to have a low heating value of 900 Btu/SCF, and the fuel oil, 136,000 Btu per gallon.)

In order to calculate the fuel requirements for a particular application, it is first necessary to assume a heat exchanger recovery. Knowing the waste gas temperature, then the temperature into the afterburner may be determined from Figure 11-8 (T_B). Entering the appropriate graph 11-10 through 11-13 at temperature T_B, and knowing the hydrocarbon concentration in the waste stream, the fuel requirement can be read directly. This does not make any allowance for heat losses which may add of the order of 5% to the heating load.

Figures 11-8 through 11-13 follow

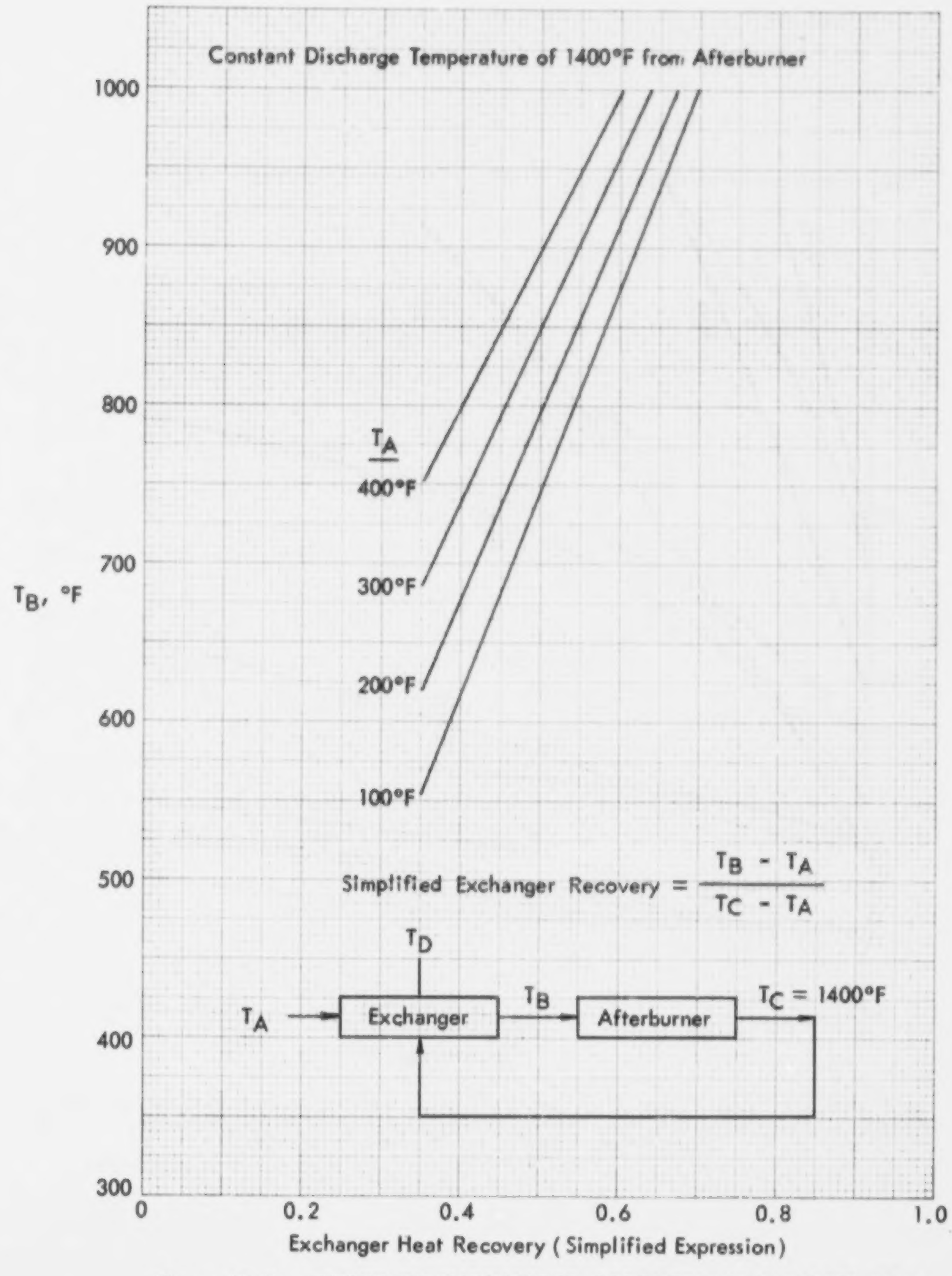

Figure 11-8. EFFECT OF EXCHANGER RECOVERY AND WASTE GAS TEMPERATURE ON INLET TEMPERATURE

Thermal Afterburners

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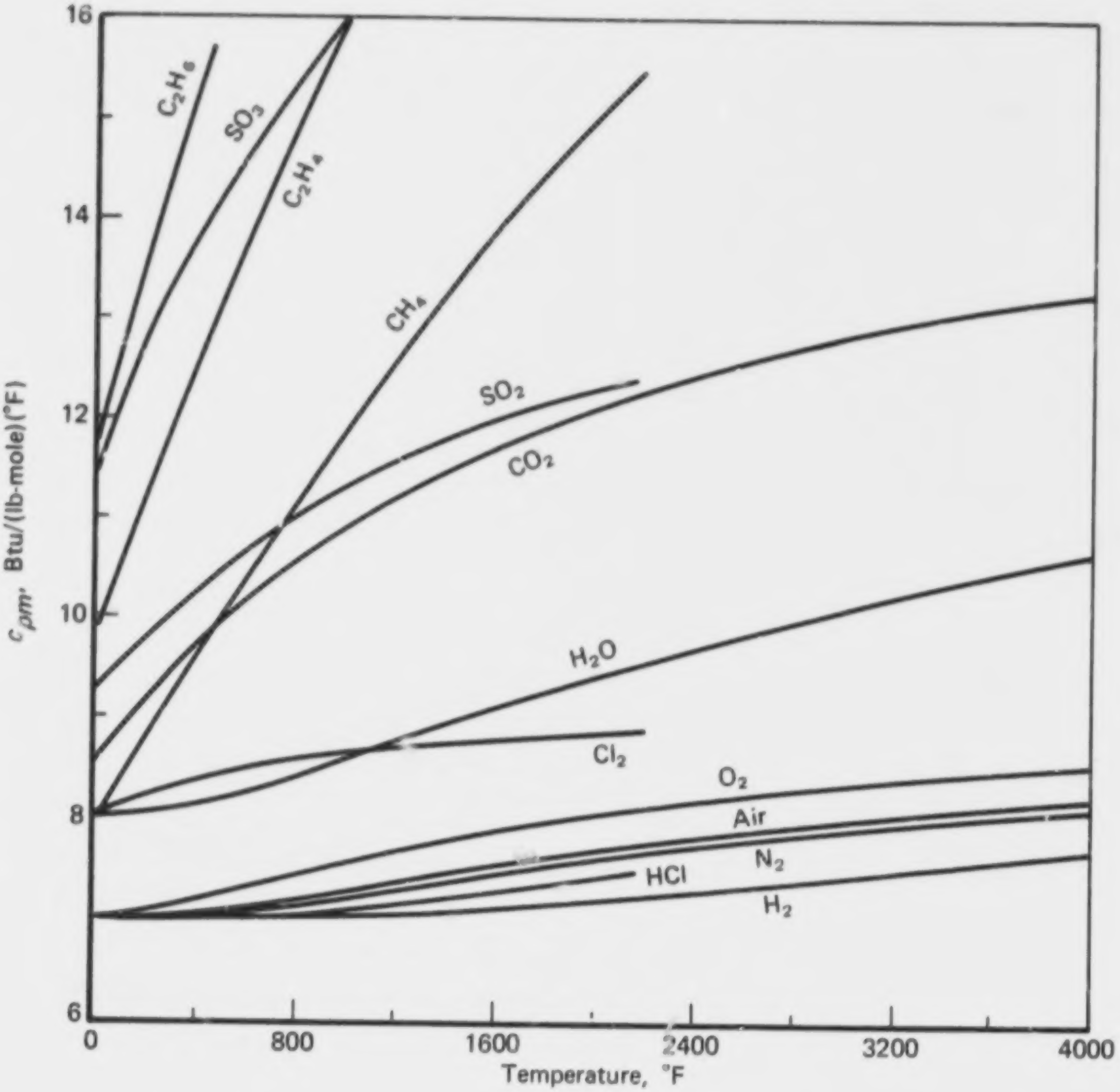

Figure 11.9. MEAN MOLAL HEAT CAPACITIES OF GASES AT CONSTANT PRESSURE
Mean Values from 77° to t°F (From References 11-1)

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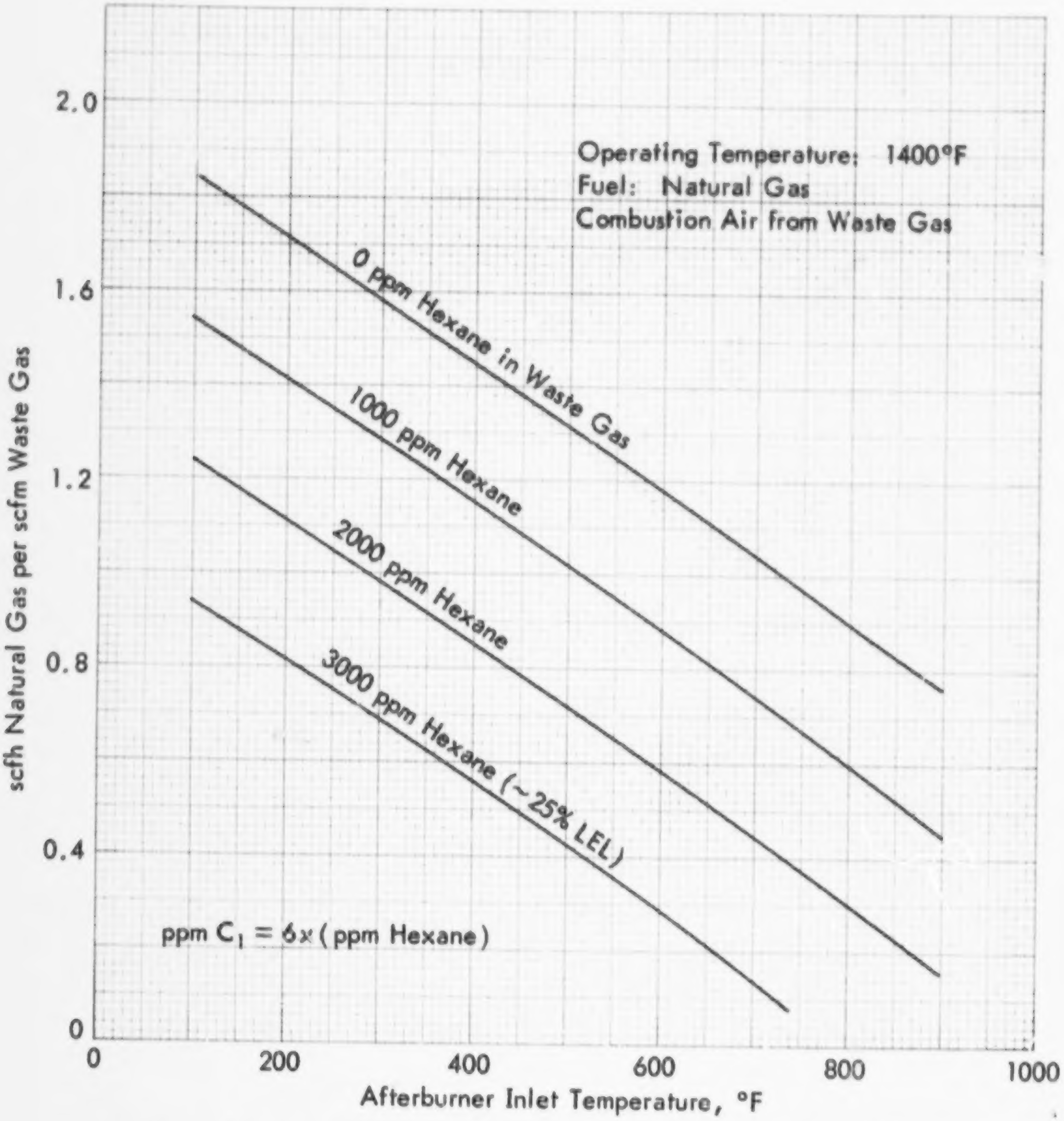

Figure 11-10. FUEL CONSUMPTION: THERMAL AFTERBURNER

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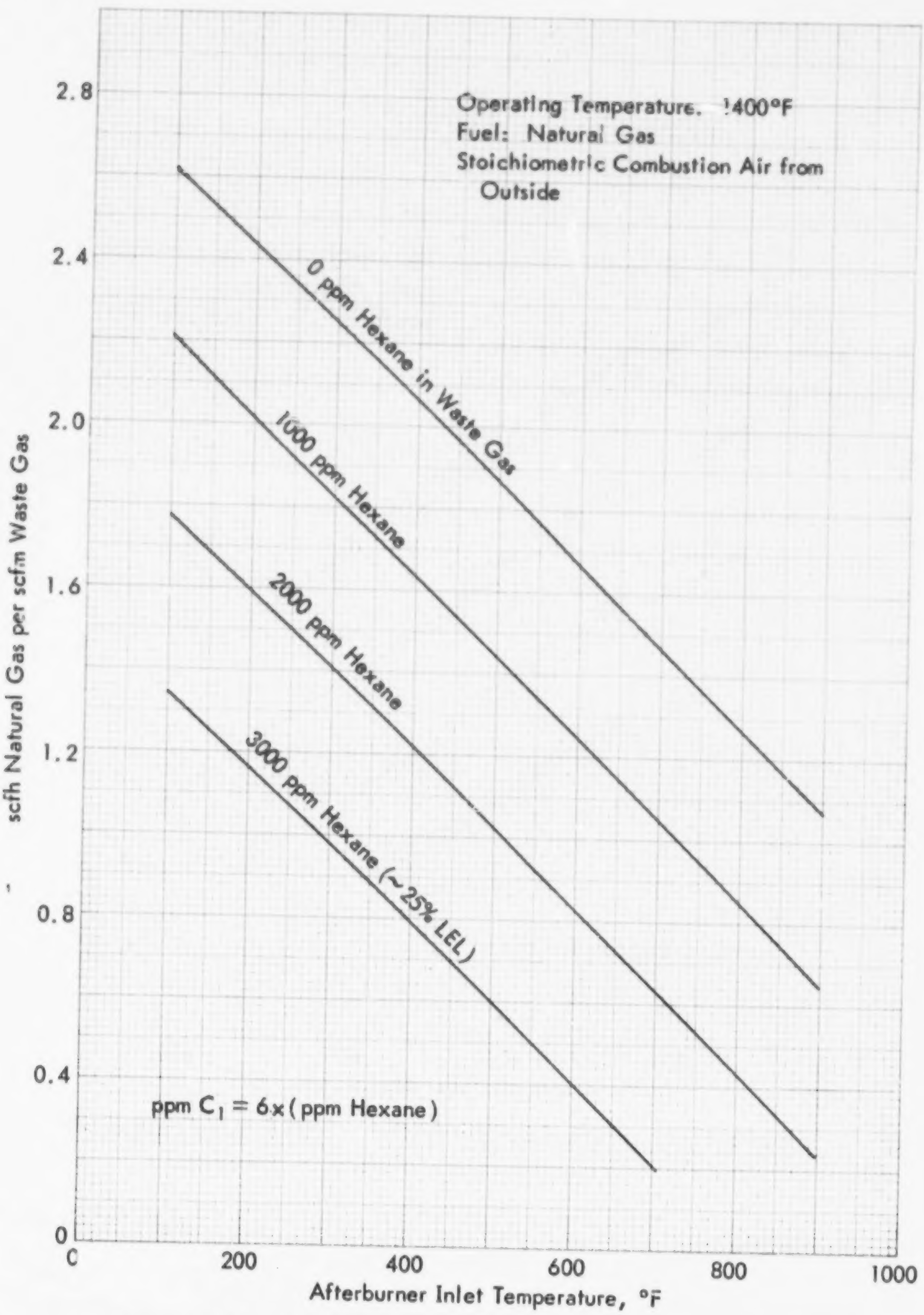

Figure 11-11. FUEL CONSUMPTION: THERMAL AFTERBURNER

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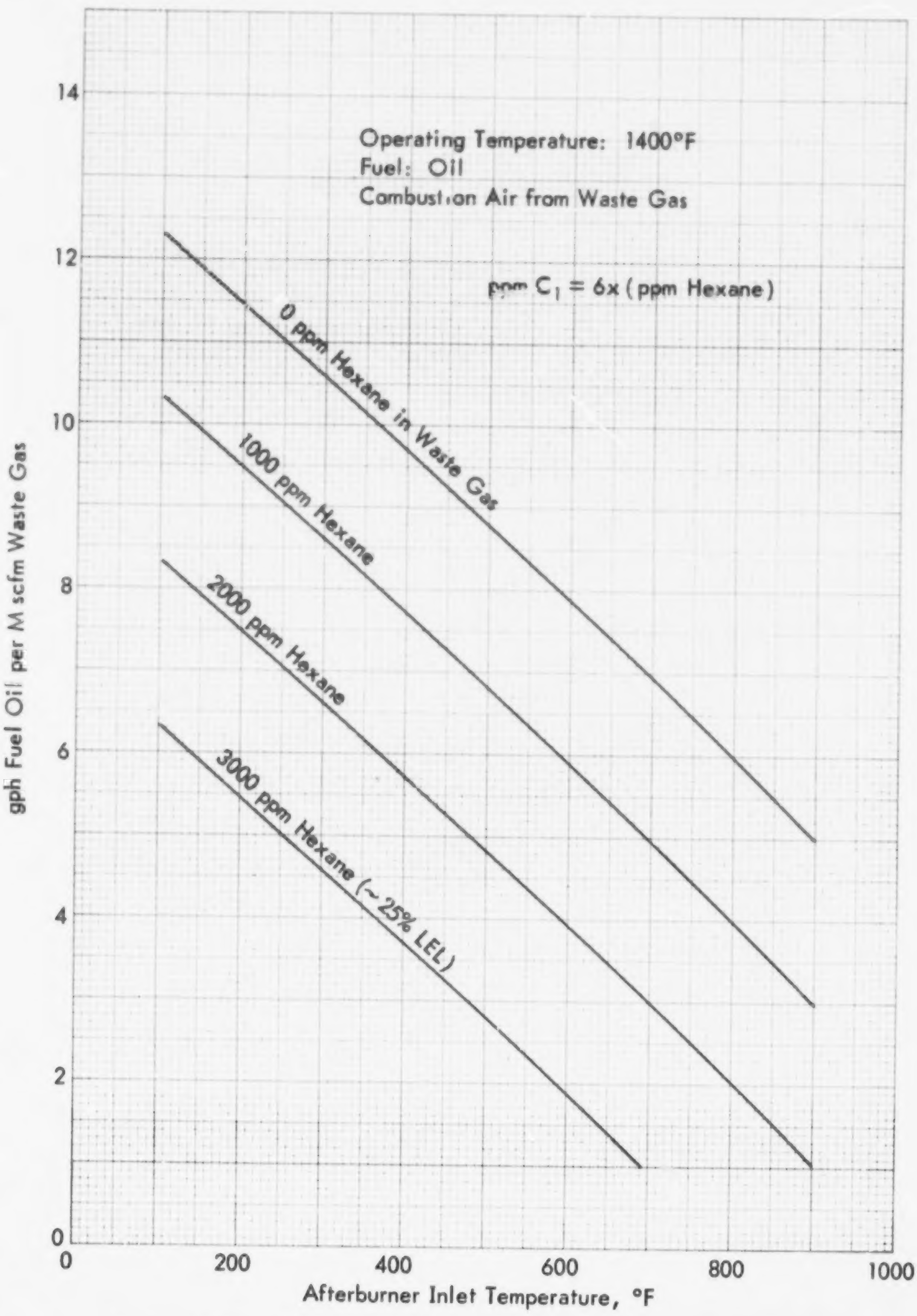

Figure 11-12. FUEL CONSUMPTION: THERMAL AFTERBURNER

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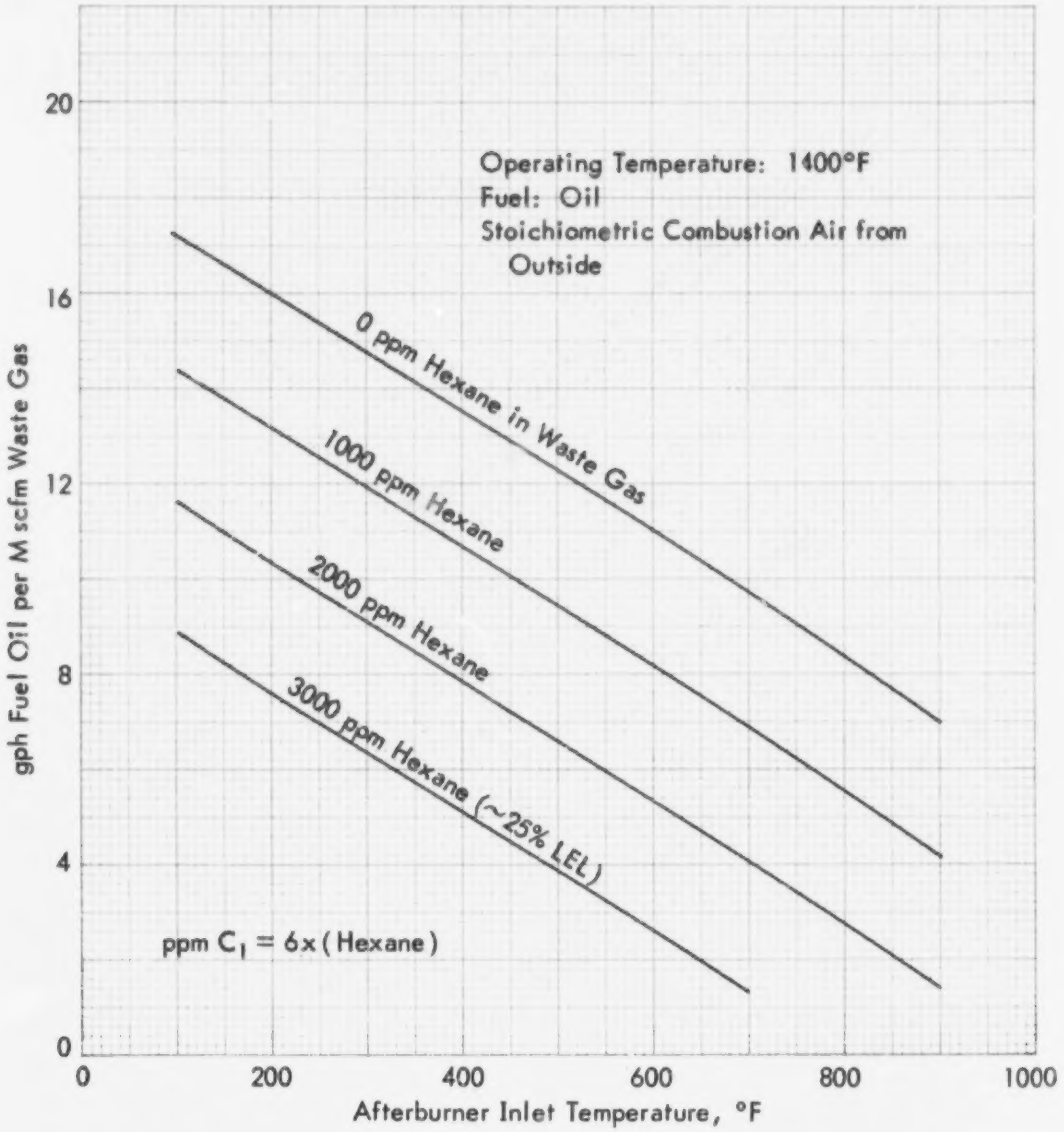

Figure 11-13. FUEL CONSUMPTION: THERMAL AFTERBURNER

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11.3 Catalytic Afterburners

11.3.1 Purchase Costs from Manufacturers' Data

Purchase costs obtained from the three major manufacturers of catalytic units have been averaged and plotted in Figure 11-14. The data have been broken into three main categories, (1) pre-engineered units, generally in smaller sizes, (2) custom designed units without heat exchangers, and (3) custom designed units with 40 - 50% heat recovery. The same information has been plotted on Figure 11-1 to show the comparison with the cost of thermal units. Pre-engineered catalytic units without heat exchangers fall in the middle range of costs for thermal units without exchangers, but the custom designed units are considerably more expensive. Catalytic units with 45% heat recovery are in the higher part of the cost range for similar thermal units. Increased capital cost of the catalytic units should be offset by fuel savings in an application for which they are suitable.

Catalysts are based on noble metals (platinum or palladium), and represent a substantial initial investment. Even though different manufacturers use different types of support for the catalyst, the average cost of the charge is approximately $1,500 for 1000 SCFM throughput. This cost has been shown on Figure 11-14 to demonstrate the large proportion of the initial cost which is tied up in catalyst alone, especially for the small pre-engineered units.

11.3.2 Instrumentation

Instrumentation requirements are generally the same as for thermal units. The waste gas is preheated prior to entering the catalyst bed with the final oxidation taking place in the bed. Fuel is controlled to maintain a constant temperature of the order of 900°F out of the catalyst bed.

11.3.3 Burner

Natural gas is used generally for preheating the waste gases although manufacturers do have designs for oil fired units. Both nozzle mix burners and distributed gas burners have been used. Again requirements are the same as for thermal units except for the lower heat input required.

11.3.4 Purchase, Installed and Operating Costs

Only 24 catalytic units were reported in replies to our questionnaires and these included two units for treating nitrogen oxides. The units were mostly old and had a low throughput. It was not felt to be worthwhile tabulating the data separately, but they are included in Table 11-2. Purchase and installed costs are not meaningful because they were installed so long ago, and in some cases because of the extensive modifications which have been made to the original equipment.

Extensive discussions were held with users of catalytic afterburners either in person or by telephone, in order to obtain as much relevant information as possible to supplement the small amount of data obtained from returned questionnaires.

Maintenance costs are largely associated with catalyst washing and replacement. In the Los Angeles area washing three times per year and annual catalyst replacement is required to maintain the level of performance demanded by the local APCD. At the other end of this scale, (not in the Los Angeles area), catalyst cleaning with an air blast every two years, and catalyst replacement every five years has been adequate. If the waste stream is free from catalyst poisons and does not contain particulate matter, over two years catalyst life would be expected with washing one to three times per year.

Washing is a relatively simple operation, but does involve removing the catalyst from the unit. In most designs the catalyst modules are small enough that they can be removed by hand. One manufacturer uses a system where the whole bed must be removed in one piece. This could be an advantage where there is adequate space available for handling, but would be a problem if the afterburner were installed in a confined space which often seems to be the case.

Catalyst replacement costs depend on the type of catalyst support being used. Where the catalyst is deposited on nichrome ribbon, it is often possible to use the same module, clean off the old catalyst, and put on new. This type of replacement cost would be of the order of $500 per MSCFM throughput. If the old module could not be re-used, replacement cost would be close to the price of a new charge at $1500 per MSCFM. Oxycat brick-type catalyst can usually be replaced for about $700/MSCFM since some of the supports can be reclaimed. This is compared to a new charge price of $1000/MSCFM in this case.

At this time manufacturers have been unable to develop a satisfactory method of reclaiming honeycomb-type supports, although they are working on the problem. Currently some manufacturers give a 10% discount for the old catalyst which represents recovery of the noble metals. Thus, replacement may be as low as $1350/MSCFM throughput. It is expected that this cost will be reduced further when techniques for reclamation are developed.

The figures quoted above are only average prices based on discussions with manufacturers. They will vary depending on the exact catalyst formulation and the manufacturer involved. It has been assumed that 1 cubic foot of catalyst is required for each 1 MSCFM throughput. (See Chapter 4 for further discussion of this.) Firm prices for catalyst replacement should be obtained directly from the manufacturer at the time the installation is being designed.

Maintenance costs, apart from catalyst replacement, would be expected to be slightly less than for the corresponding thermal units because of the lower operating temperature.

Operating labor is again a negligible factor.

Fuel costs are considerably lower than for thermal units because of the lower operating temperature.

Figures 11-14 and 11-15 follow

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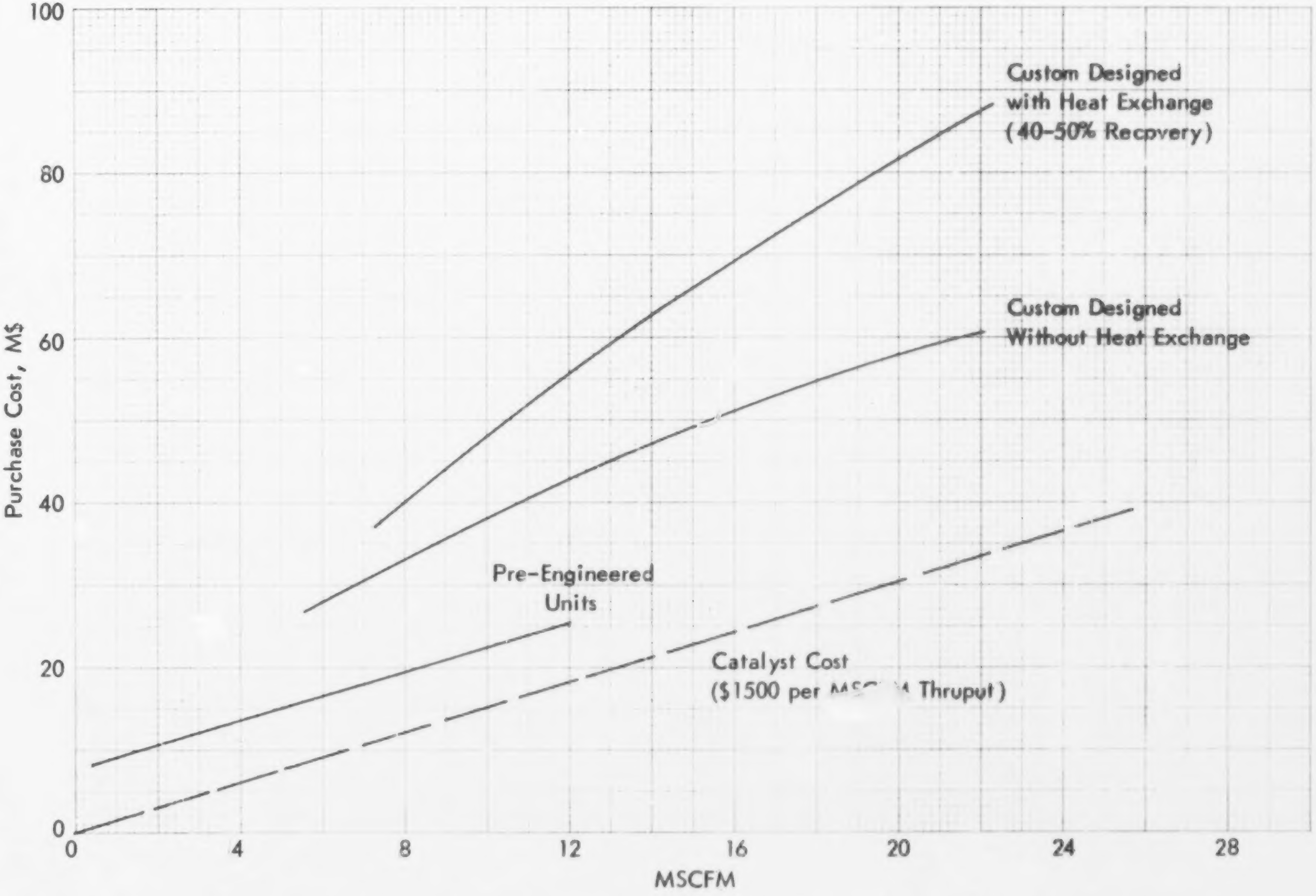

Figure 11-14. CATALYTIC AFTERBURNER COSTS BASED ON MANUFACTURERS' INFORMATION

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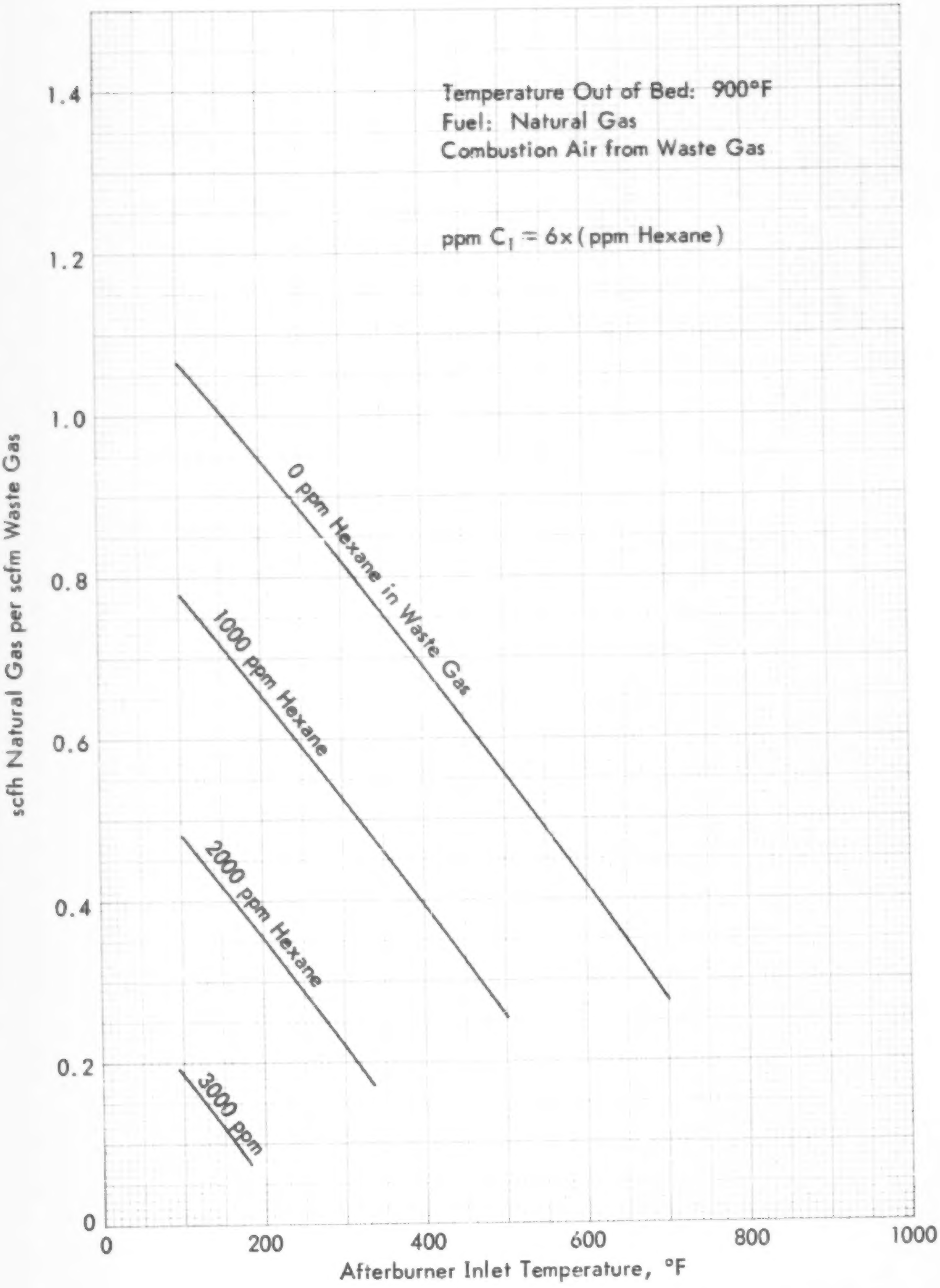

Figure 11-15. FUEL CONSUMPTION: CATALYTIC AFTERBURNER

11.3.5 Fuel

Catalytic afterburners should give satisfactory hydrocarbon reduction at an outlet temperature of 900°F from the catalyst bed (see Chapter 4). Part of the temperature rise is provided by fuel in the preheat section, and part by oxidation in the catalyst bed.

Fuel requirements have been calculated and presented graphically in Figure 11-15 through 11-18 for the 900°F discharge temperature. They include both natural gas and fuel oil with combustion air either from the waste gas or from the outside with different levels of hydrocarbon concentration in the waste gas. Figure 11-19 shows the effect of heat exchanger recovery on the temperature rise through the exchanger.

Determination of fuel requirements follows the same procedure as for thermal afterburners. An exchanger heat recovery is assumed, then from the waste gas temperature T_A, the temperature into the afterburner T_B, may be read from Figure 11-19. Entering the appropriate graph 11-15 through 11-18 at temperature T_B, and knowing the hydrocarbon concentration in the waste stream, the fuel requirement can be read directly. These numbers are average values, and somewhat lower fuel usage would be expected with a new catalyst bed, and somewhat higher as the catalyst ages. Generally, catalyst on a ceramic base is more active than catalyst on a nichrome ribbon base, so the former would give slightly lower fuel consumption. An allowance for heat losses of the order of 5% should be added.

11.4 Total Annual Cost

The total annual cost of operating an afterburner depends on the installed cost, depreciation, interest charges, as well as the maintenance, fuel and power costs. All these factors must be considered when an installation is being planned, and must be based on firm quotations from equipment suppliers coupled with the local costs for fuel, power, maintenance labor and construction.

Illustrative examples of the cost of catalytic units and thermal units with and without heat exchangers have been calculated for comparison, and the results are given in Table 11-2. It must be emphasized that these do not represent real costs since factors on capital have been used to estimate installation and maintenance costs which will vary depending on the unit and the plant location. The following assumptions were made:

1) Waste stream 100°F, 10,000 SCFM (= 10,800 ACFM at 100°F), containing 1000 ppm hexane;

2) Combustion air provided by the waste gas;

3) Installed cost 2X purchase cost;

4) Ten-year life straight-line depreciation;

5) Interest charge on capital, 9%;

6) Catalyst replacement every two years;

7) Maintenance 2% of installed cost;

8) Fuel natural gas 50c/1000 SCF. Requirements calculated using Figures 11-8, 11-10, 11-15, and 11-19;

9) Blower power calculated assuming forced draft, 6 in. and 10 in. pressure drop for catalytic units with and without exchangers; 2 in., 6 in., and 10 in. for thermal units without exchanger, with 40% heat recovery and with 60% heat recovery, respectively.

10) Electricity 1c/KW hr.;

11) 4000 hours/year operation;

12) Individual costs have been rounded off;

13) One ft^3 of catalyst/1000 SCFM waste gas.

Using this particular set of assumptions , it is seen that the pre-engineered catalytic unit is somewhat cheaper than all the others, followed by the catalytic unit with heat exchanger, and then by the thermal unit with a 40% exchanger. A thermal unit with a 60% exchanger is the most expensive, but this cost is heavily influenced by using straight percentage factors for installation and maintenance.

Figures 11-16 through 11-19 follow
Tables 11-1 and 11-2 follow

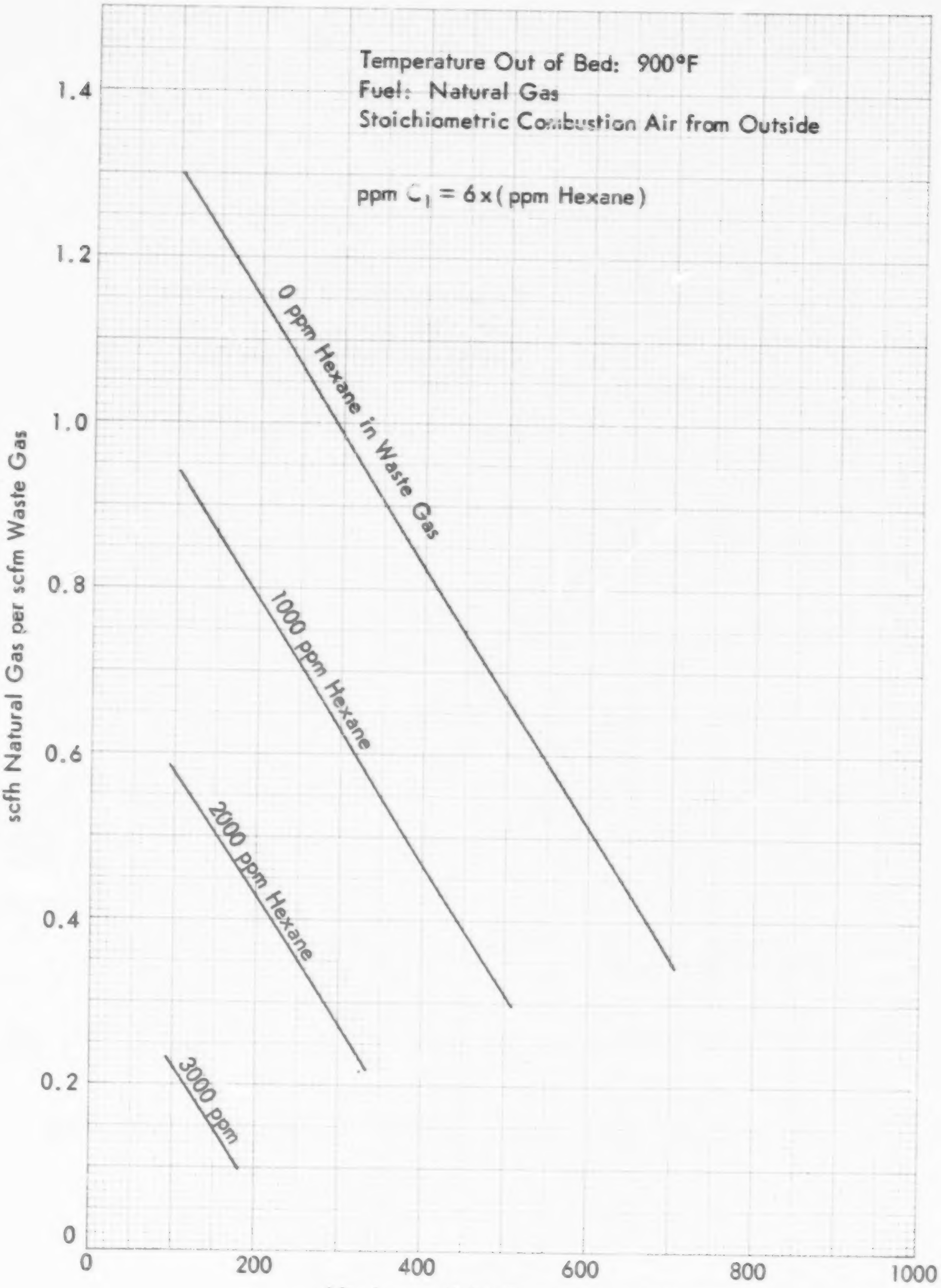

Figure 11-16. FUEL CONSUMPTION: CATALYTIC AFTERBURNER

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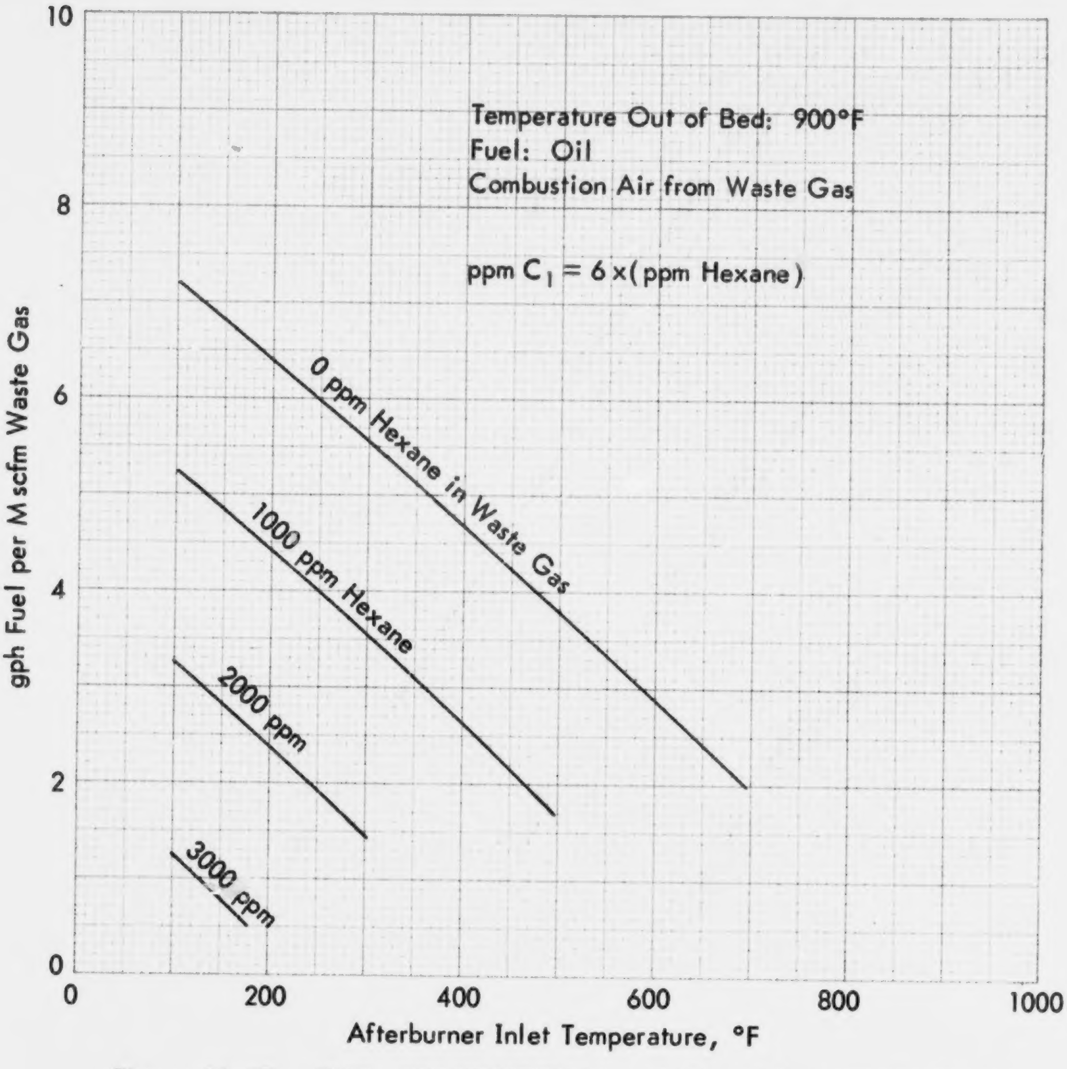

Figure 11-17. FUEL CONSUMPTION: CATALYTIC AFTERBURNER

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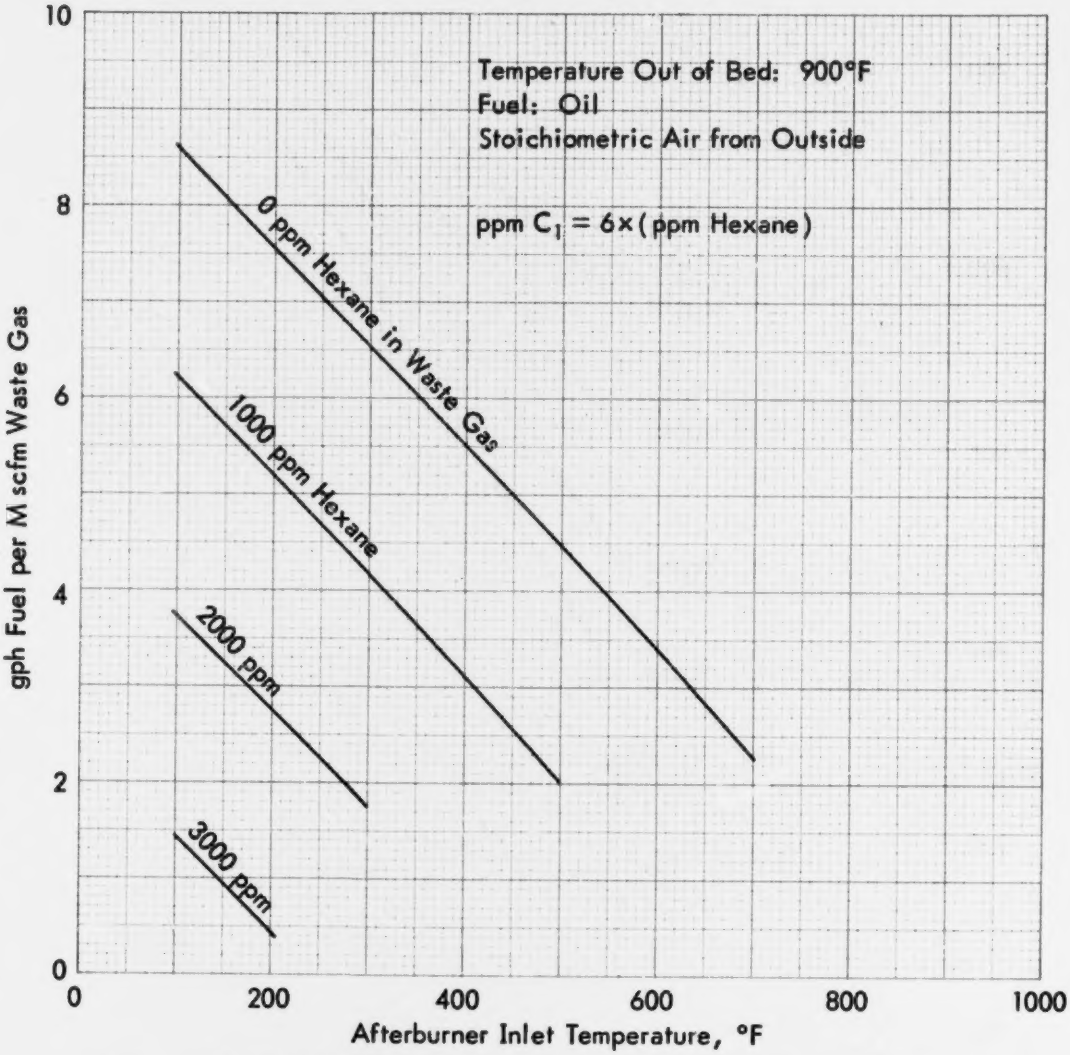

Figure 11-18. FUEL CONSUMPTION: CATALYTIC AFTERBURNER

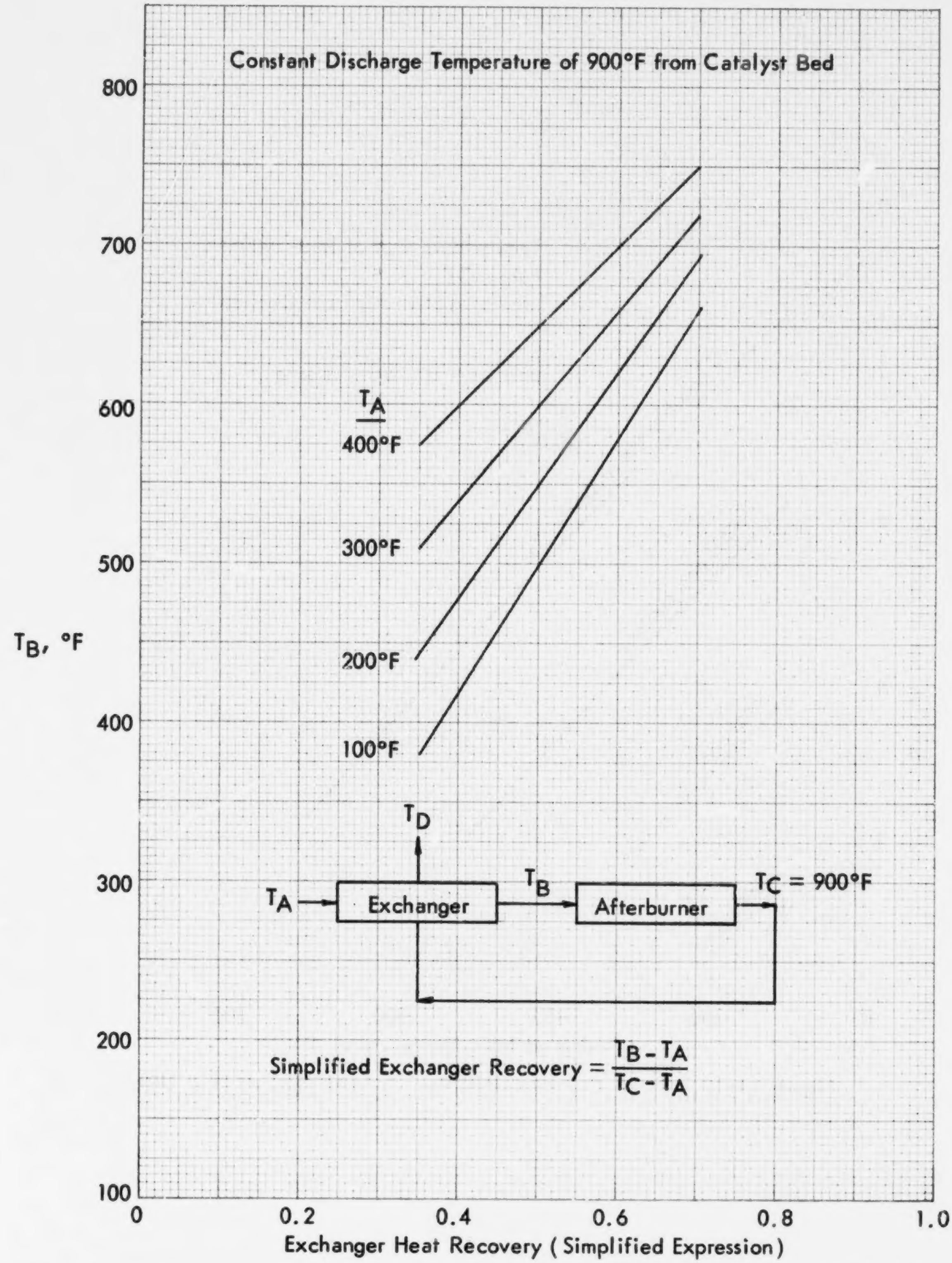

Figure 11-19. EFFECT OF EXCHANGER RECOVERY AND WASTE GAS TEMPERATURE ON INLET TEMPERATURE

Catalytic Afterburners

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Table 11-1.

							WASTE STREAM			CONTAMINANT			AFTERBURNER											FUEL	COST M$				
NO	PROCESS CATEGORY	PERFORMANCE	FUME	THERM/CAT	PERCENT AVILABLE	FUEL	MSCFM	TEMP	PERCENT OXYGEN	LBS PER HOUR	PPM	BTU PER SCF	BATCH/CONT	HOURS/DAY	PRETREAT	HEATEXCH	IND/FORCED	TEMP IN	TEMP OUT	EQ DIA	LENGTH	ORG OUT	RECYCLE	MM BTU HOUR	PURCH	INSTD	OP PER YEAR	PROBLEMS	YR
26A	8061	N	D	T	99	G							C	15	N	N			1200	5.5	6.0		N						69
26B	8061	N	D	T	99	G							C	15	N	N			1200	5.5	6.0		N						69
91	4952	S	A	T	95	Z	27.0	75					C	24	N	Y	F	1200	1450					30.0					63
92	4952	S	A	T	95	G	10.0	75					C	24	N	Y			1450	6.5	5.5			2.5					63
93	4952	S	A	T	95	G	15.0	75					C	24	N	N		75	1450	8.5	4.5			6.0					63
36	3900	S	G	T	99	G	65.0	200	18				C	5	N	N	F	200	1500	8.0	25.0		N	136.0		560.0	60.0		68
41	3900	S	D	T	99	G	3.0	70		5			C	24	N	N	I	70	1600				N	.5	6.0	13.0	4.0		65
56	3900	S	C	T	90	G	7.4	70				455.0	B	1		N	F	70	1250	9.0	22.0		N	76.0	48.9	50.0	.5	REFRACTORY	67
69	3900	S	D	T	50	G		1000					B	3	N	N	F	1000	1450	5.0	7.0		N	.5	5.2	10.0		FLAME ROD	71
77	3900	S	A	T	99	Z									N	N	I	1200	1600	7.6	12.0		N						69
82A	3900	S	G	T	98	G	10.0	350		5			C	24	N	N	F	325	1350	6.0	20.0	R95	N	9.7	11.0	65.0	90.0		70
82B	3900	S	G	T	98	G	10.0	350		5			C	24	N	N	F	325	1350	6.0	20.0	R95	N	9.7	11.0	65.0	90.0		70
10	3714	S	G	T		G							B	9	N	N	I			3.4	6.7		N		4.0		1.7		62
57	3691	S	E	T	99	G	.4	75		50	10000		C	7	N	N	F	75	1900	1.5	35.0	1	N	3.2	21.0	29.0	6.0	NOX	70
75	3691	S	D	T	95	Z	.3	100			310	.7	C	24	N	N	F	100	1000	2.0	8.0		N		5.0	8.0		FLAME ROD FAIL	70
80	3621	S	B	T	99	G	1.0	440					B	5	N	N	F	440	1300	3.0	4.0		N	1.4	4.6	7.1			71
83A	3621	S	D	T	99	G		400					C	7	N	N		400	1400	1.7	4.0		N		5.0	6.0			70
83B	3621	S	D	T	99	G		400					C	7	N	N		400	1400	1.7	4.0		N		5.0	6.0			70
1	3479	S	B	T	99	G	5.5	250	20	140			C	24	N	Y	I	400	1400	5.0	5.0	270	N			25.0			
5A	3479	S	B	T	99	G	3.4	250	18	50	2300	5.0	C	18	N	N	F	250	1400	2.7	21.0	123	N	3.4	6.0	12.0	7.0	SAFETY DEVICES	62
5B	3479	S	B	T	99	G	7.7	230	14	290	6000	14.0	C	18	N	N	F	230	1375	2.9	91.0	117	Y	4.5	30.0	60.0	3.0	SAFETY DEVICES	68
5C	3479	S	B	T	99	G	3.4	250	14	50	2300	5.0	C	18	N	N	F	250	1300	2.4	80.0	123	Y	3.4	16.0	32.0	1.0	SAFETY DEVICES	65
5D	3479	S	B	T	99	G	3.4	250	14	50	2300	5.0	C	18	N	N	F	250	1300	2.4	80.0	123	Y	3.4	16.0	32.0	1.0	SAFETY DEVICES	66
5E	3479	S	B	T	99	G	6.5	250	18	260	2300	10.0	C	18	N	N	F	250	1400	3.5	25.0	125	N	4.5	16.5	19.0	10.0	SAFETY DEVICES	60
5F	3479	S	B	T	99	G	3.8	230	18	50	2300	5.0	C	13	N	N	F	230	1380	2.6	21.0	125	Y	3.4	9.0	18.0	7.0	SAFETY DEVICES	62
5G	3479	S	B	T	99	G	3.8	230	18	50	2300	5.0	C	13	N	N	F	230	1380	2.6	21.0	125	Y	3.4	9.0	18.0	7.0	SAFETY DEVICES	63
5H	3479	S	B	T	99	G	3.4	250	18	50		5.0	C	18	N	N	F	250	1400	2.9	91.0		Y	2.2	26.0	50.0	1.3	SAFETY DEVICES	67
5I	3479	S	B	T	99	G	3.4	250	18	50		5.0	C	18	N	N	F	250	1400	2.9	91.0		Y	2.2	26.0	50.0	1.3	SAFETY DEVICES	68
5J	3479	S	B	T	99	G	8.2	300	14	275	6000	11.0	B	18	N	N	F	250	1250	3.6	77.0	117	Y	6.0	23.0	45.0	3.5	SAFETY DEVICES	67
5K	3479	S	B	T	98	G	1.0	300	18	5	1300	2.0	C	18	N	N	I	300	1430	2.3	4.3	100	N	1.5	6.0	12.0	3.0	BAFFLE FAILED	68
5L	3479	S	B	T	98	G	1.0	300	18	5	1300	2.0	C	18	N	N	I	300	1430	2.3	4.3	100	N	1.5	6.0	12.0	3.0	BAFFLE FAILED	68
5M	3479	S	B	T	98	G	1.0	300	18	5	1300	2.0	C	18	N	N	I	300	1430	2.3	4.3	100	N	1.5	6.0	12.0	3.0	BAFFLE FAILED	68
5N	3479	S	B	T	98	G	1.0	300	18	5	1300	2.0	C	18	N	N	I	300	1430	2.3	4.3	100	N	1.5	6.0	12.0	3.0	BAFFLE FAILED	68
5O	3479	S	B	T	98	G	1.0	300	18	5	1300	2.0	C	18	N	N	I	300	1430	2.3	4.3	100	N	1.5	6.0	12.0	3.0	BAFFLE FAILED	68
6	3479	S	D	T		G	1.0	260	10		150		C	16	N	N	I	260	1400				N		10.0	12.0	6.3	BAFFLE FAILED	67
8	3479	S	B	T	96	G					600		C	24	N	Y	F		1400			100	Y			50.0		BRICKS LOOSE	
9A	3479	S	B	T	99	G	9.5	250	21	137	826		C	8	N	N	I	250	1350	5.5	7.3	80	N	11.3	15.0	52.0	14.0	REFRACTORY	67
9B	3479	N	B	T	99	G	13.0	250	21	86	438		C	8	N	N	I	250	1350	5.5	9.0	R86	N	15.4	18.0	80.0	19.0	REFRACTORY	67
9C	3479	S	B	T	99	G	10.0	250	21	102	478		C	8	N	N	I	250	1350	5.5	6.9	95	N	11.9	19.0	72.0	15.0	REFRACTORY	67
9D	3479	N	B	T	99	G	6.7	250	21	84	590		C	8	N	N	I	250	1350	5.0	6.4	R83	N	9.1	11.0	57.0	10.1	REFRACTORY	67
9E	3479	N	B	T	99	G	5.7	250	21	93	427		C	8	N	N	I	250	1350	4.5	7.0		N	6.6	11.0	40.0	8.7	REFRACTORY	67
11	3479	N	B	T		G	.8	320		1	770		C	10	N	N	I	330	1450	2.8	9.5	170	Y		9.0	29.0		FAILS CODE	69
22	3479	S	B	T	98	G	1.5	350	10				C	8	N	N	I	325	1200	2.0	8.6		N	1.7		7.0		FAN CONTRL REFR	
32A	3479	S	B	C	99	G	1.7	425			2000		C		N	N	I	425					N			20.0			68
32B	3479	S	D	T	99	G	3.0	100					C	24	N	N	I	100	1200	3.3	18.0		N	6.9		8.0			67
32C	3479	S	B	T		G	5.5	220			2200		C	16	N	N	F	220	1200	4.5	6.0		N	19.0	11.0	19.0			69
32D	3479	S	B	T		G	1.4	250			2200		C	2	N	N	F		1400	2.0	8.0		N			8.6			69
32E	3479	S	B	T		G	3.1	250			2200		C	16	N	N	F		1100	3.5	15.0		N			9.1			68
32F	3479	S	B	T		G	7.2	250			2200		C	8	N	N	F		1400	4.5	15.0		N			20.0			68
32G	3479	S	B	T		G	3.2	250			2200		C		N	N	F		1400	2.2	15.0		N			8.3			68
35A	3479	S	G	T		G							B	4	D	N	F			1.6	12.0			.5		5.5		FLAME ROD POSN.	67

PROCESS CATEGORY FROM US GOVERNMENT STANDARD INDUSTRIAL CLASSIFICATION
FUME A=ODOR B=HC REDUCTION C=TOXIC FUMES D= SMOKE/PARTICULATES E=OTHER F=A+B G=B+D H=A+D I=A+B+D
FUEL G=NATURAL GAS O=OIL M=EITHER O OR G Z=OTHER PPM = PARTS PER MILLION CI
ORGANICS OUT EITHER PARTS PER MILLION OR R90=90 PERCENT REDUCTION THRU AFTERBURNER
PERFORMANCE S=SATISFACTORY N=NOT SATISFACTORY X=NOT YET RUN
DIMENSIONS IN FEET TEMPERATURES DEGREES F IND/FORCED=INDUCED OR FORCED DRAFT N=NATURAL DRAFT
PRETREAT N=NONE C=CYCLONE D=DILUTION S=SCRUBBER R=ROTOCLONE E=PRECIPITATOR PROBLEMS OS=OUT OF SERVICE

Table 11-1. (Contd-1.)

								********WASTE STREAM*********						********************AFTERBURNER********************																
NO	PROCESS CATEGORY	PERFORMANCE	FUME	THERM/CAT	PERCENT AVAILABLE	FUEL	MSCFM	TEMP	PERCENT OXYGEN	CONTAMINNT LBS PER HOUR	PPM	BTU PER SCF	BATCH/CONT	HOURS/DAY	PRETREAT	HEAT EXCH	IND/FORCED	TEMP IN	TEMP OUT	EQ DIA	LENGTH	ORG OUT	RECYCLE	FUEL MM BTU HOUR	****COST M$***** PURCH	INSTD	OP PER YEAR	****PROBLEMS***	YR
35B	3479	S	G	T		G							B	4	D	N	F			5.1	17.0			1.3		6.6		FLAME ROD POSN.	67
37A	3479		B	T		G	2.0								N		I		1200				N			17.0			71
45A	3479	S	B	T	90	G	10.0	300	20	125	4000	125.0	C	20	N	Y	F	800	1400	6.0	12.0	300	N	4.0	45.0	60.0	15.0	FAN HT XCHANGER	68
45B	3479	S	B	T	90	G	10.0	300	20	125	4000	125.0	C	20	N	Y	F	800	1400	6.0	12.0	300	N	4.0	45.0	60.0	15.0	FAN HT XCHANGER	68
45C	3479	S	B	T	90	G	10.0	300	20	125	4000	125.0	C	20	N	Y	F	800	1400	6.0	12.0	300	N	4.0	45.0	60.0	15.0	FAN HT XCHANGER	68
45D	3479	S	B	T	90	G	10.0	300	20	125	4000	125.0	C	20	N	Y	F	800	1400	6.0	12.0	300	N	4.0	45.0	60.0	15.0	FAN HT XCHANGER	68
45E	3479	S	B	T	90	G	10.0	300	20	125	4000	125.0	C	20	N	Y	F	800	1400	6.0	12.0	300	N	4.0	45.0	60.0	15.0	FAN HT XCHANGER	68
50A	3479	S	B	T	99	G	5.0	350					C	20	N	N	I	350	1300	4.0	10.0		N		20.0	75.0	11.0		68
50B	3479	S	B	T	99	G	7.5	350					C	20	N	Y	F	850	1300	5.0	15.0		N		20.0	75.0	14.5	UNSATISFACTORY	67
50C	3479	X	B	T	99	G	9.0	350					C	20	N	Y	I	850	1300	4.5	4.0		N		57.0	100.0	14.5		71
54A	3479	S	B	T	99	G	8.3	240		18	1170		C	16	N	Y	I	705	1400	5.5	10.0	35	Y	7.5		80.0	17.0	MODS AFTER INST	67
54B	3479	S	B	T	99	G	9.6	260		17	1233		C	16	N	Y	I	680	1400	5.5	10.0	52	N	6.1		90.0	9.8	MODS AFTER INST	67
54C	3479	S	B	T	99	G	4.5	200		1	142		C	16	N	Y	I	620	1400	5.3	8.0	6	Y	5.7		63.0	11.5	MODS AFTER INST	67
54D	3479	S	B	T	99	G	6.7	230		3	402		C	16	N	Y	I	640	1400	5.4	8.0	4	N	5.1		80.0	11.0	MODS AFTER INST	67
54E	3479	S	B	T	99	G	1.9	290		1	283		C	16	N	Y	I	600	1400	2.9	6.0	22	N	.9		30.0	9.3	MODS AFTER INST	67
54F	3479	S	B	T	99	G	8.1	200		14	126		C	16	N	Y	I	620	1400	5.5	10.0	9	Y	6.2		72.0	16.0	MODS AFTER INST	67
54G	3479	S	B	T	99	G	5.2	90		3	430		C	16	N	Y	I	720	1400	3.8	10.5	36	N	3.3		39.0	11.3	MODS AFTER INST	67
59A	3479	S	A	C	98	G	1.9	300		80			C	10	N	N	F	450	900				N	.2	7.2	9.7	2.2	CAT POISON	61
59B	3479	S	A	C	98	G	1.9	300		80			C	10	N	N	F	450	900				N	.2	7.2	9.7	2.2	CAT POISON	61
59C	3479	S	A	T	99	G	1.9	350		80			C	16	N	N	F	350	900	3.4	14.5		N	2.2	16.5	19.0	5.3		71
60	3479	X	H	T		G	8.8	275	22	93			C	18	N	Y	F	900	1500				N	14.0	40.5	60.0	6.7		71
62	3479	S	B	C	99	G	1.5	300		40			C	18	N	N	I	800	1000				N	1.0	11.5	13.5	3.5		65
63	3479	S	B	C	99	G	1.4	300		40			C	18	N	N	I	800	1000				N	1.0	7.5	8.5	4.0		59
71A	3479	X	B	T		G	8.9	180		90		5.7	C	20	N	Y	F	980	1400	3.9	16.0		N		17.0	38.0		NOT OPERATG YET	71
71B	3479	X	B	T		G	8.9	180		90		5.7	C	20	N	Y	F	980	1400	3.9	16.0		N		33.0	77.0		NOT OPERATG YET	71
72	3479	S	B	T	99	G	7.9	295		90		5.7	C	20	N	Y	I	655	1400	4.0	20.0		N	1.7		56.0	12.0	REFRACTORY FAIL	70
74	3479	S	B	T	90	G	3.9	400			4000		C	16	N	N	I	900	1500	5.3	16.8	5	N	11.1	20.0	30.0	40.0	TEMP CONTROL	68
79A	3479	S	B	T		G	15.5	270			2822		C	16	N	N	F	270	1400	7.5	33.0	12	N	15.0	35.0	50.0	50.0		67
79B	3479	S	B	T	99	G	12.5	350			965		C	16	N	Y	F	814	1400	4.4	8.4	4	N	8.0	50.0	75.0	18.0	HT EX REPLACED	68
79C	3479	S	B	T	99	G	11.0	350			1276		C	16	N	Y	F	803	1400	4.6	8.9	4	N	7.8	50.0	75.0	13.0	HT EX REPLACED	68
79D	3479	S	B	T	99	G	7.9	350			409		C	16	N	Y	F	800	1400	4.3	7.6	5	N	5.6	50.0	75.0	10.0	HT EX REPLACED	68
79E	3479	S	B	T	99	G	4.2	350			185		C	16	N	Y	F	795	1400	3.0	7.2	7	N	3.0	50.0	75.0	6.0	HT EX REPLACED	68
79F	3479	S	B	T	99	G	4.7	350			536		C	16	N	Y	F	808	1400	3.1	6.8	4	N	3.3	50.0	75.0	6.9	HT EX REPLACED	68
85A	3479	N	B	C	99	G	3.9	275		36	3000		C	24	N	Y	F	700	1050	7.6	8.5	300	N	3.2				CATALYST LIFE	59
85B	3479	N	B	C	99	G	4.1	275		36			C	24	N	N	F	275	1050	8.1	8.5		N	4.2				CATALYST LIFE	59
85C	3479	N	B	C	99	G	6.5	275		36			C	24	N	Y	F	700	1050	7.6	10.5		N	5.1				CATALYST LIFE	59
86A	3479	S	B	T	99	G	10.0	275		100			C	24	N	Y		620	1400				N	7.8				EXCH. FAILS	
86B	3479	S	B	T	99	G	10.0	275		100			C	24	N	Y		620	1400				N	7.8					
86C	3479	S	B	T	99	G	10.0	275		100			C	24	N	Y		620	1400				N	7.8					
86D	3479	S	B	T	99	G	10.0	275		100			C	24	N	Y		620	1400				N	7.8					
86E	3479	S	B	T	99	G	7.4	275		36			C	24	N	Y		620	1400				N	6.5					
86F	3479	S	B	T	99	G	7.4	275		36			C	24	N	Y		620	1400				N	6.5					
87A	3479	S	B	T	99	G	5.0	275					C	24	N	Y	F	850	1400	4.0	20.0		N	4.0					62
87B	3479	S	B	T	99	G	5.0	275					C	24	N	Y	F	850	1400	4.0	20.0		N	4.0					62
87C	3479	S	B	T	99	G	5.0	275					C	24	N	Y	F	850	1400	4.0	20.0		N	4.0					62
87D	3479	S	B	T	99	G	10.0	275					C	24	N	Y	F	850	1400	6.0	40.0		N	8.0		100.0			69
87E	3479	S	B	T	99	G	10.0	275					C	24	N	Y	F	850	1400	6.0	40.0		Y	8.0		100.0			69
88A	3479	N	B	C	99	G	5.2	275		100			C	24	N	Y	F	700	1000	7.6	10.5		N	2.0				CATALYST LIFE	59
88B	3479	N	B	C	99	G	5.6	275		100			C	24	N	Y	F	600	1000	7.6	10.5		N	2.0				CATALYST LIFE	55
89A	3479	S	B	T	99	G	10.0	275					C	24	N	Y	F	850	1400	4.0	20.0			8.0					62
89B	3479	S	B	T	99	G	12.0	275					C	24	N	Y	F	850	1400	5.0	30.0			10.0					

PROCESS CATEGORY FROM US GOVERNMENT STANDARD INDUSTRIAL CLASSIFICATION
FUME A=ODOR B=HC REDUCTION C=TOXIC FUMES D= SMOKE/PARTICULATES E=OTHER F=A+B G=B+D H=A+D I=A+B+D
FUEL G=NATURAL GAS O=OIL M=EITHER O OR G Z=OTHER PPM = PARTS PER MILLION C1
ORGANICS OUT EITHER PARTS PER MILLION OR R90=90 PERCENT REDUCTION THRU AFTERBURNER
PERFORMANCE S=SATISFACTORY N=NOT SATISFACTORY X=NOT YET RUN
DIMENSIONS IN FEET TEMPERATURES DEGREES F IND/FORCED=INDUCED OR FORCED DRAFT N=NATURAL DRAFT
PRETREAT N=NONE C=CYCLONE D=DILUTION S=SCRUBBER R=ROTOCLONE E=PRECIPITATOR PROBLEMS OS=OUT OF SERVICE

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Table 11-1. (Contd-2.)

NO	PROCESS CATEGORY	PERFORMANCE	FUME	THERM/CAT	PERCENT AVAILABLE	FUEL	********WASTE STREAM******** MSCFM	TEMP	PERCENT OXYGEN	CONTAMINANT LBS PER HOUR	PPM	BTU PER SCF	********AFTERBURNER******** BATCH/CONT	HOURS/DAY	PRETREAT	HEAT EXCH	IND/FORCED	TEMP IN	TEMP OUT	EQ DIA	LENGTH	ORG OUT	RECYCLE	FUEL MM BTU HOUR	****COST M$**** PURCH	INSTD	OP PER YEAR	****PROBLEMS****	YR
89C	3479	S	B	T	99	G	12.0	275					C	24	N	Y	F	850	1400	5.0	30.0			10.0					
89D	3479	S	B	T	99	G	19.0	275					C	24	N	Y	F	850	1400	6.0	40.0			15.0					67
90A	3479	N	B	T	99	G	10.0	275					C	24	N	N	F	275	1400	5.9	11.0			3.8		21.0	9.0	ODOR HC IN DIS.	
90B	3479	N	B	T	99	G	7.0	275					C	24	N	N	F	275	1400	5.9	11.0			3.8		21.0	8.0	ODOR HC IN DIS.	
90C	3479	N	B	T	99	G	7.0	275					C	24	N	N	F	275	1400	5.9	11.0			2.7		21.0	8.0	ODOR HC IN DIS.	
90D	3479	N	B	T	99	G	7.0	275					C	24	N	N	F	275	1400	5.9	11.0			2.7		21.0	8.0	ODOR HC IN DIS.	
90E	3479	N	B	T	99	G	7.0	275					C	24	N	N	F	275	1400	5.9	11.0			2.7		21.0	8.0	ODOR HC IN DIS.	
90F	3479	N	B	T	99	G	7.0	275					C	24	N	N	F	275	1400	5.9	11.0			2.7		21.0	8.0	ODOR HC IN DIS.	
97A	3479	S	B	T	99	G	10.0	280			650		C	8	N	Y	F	600	1400			55	N		63.0		8.3		68
97B	3479	S	B	T	99	G	12.0	235			650		C	8	N	Y	F	600	1400			55	N		63.0		8.3		68
98A	3479	S	B	T	99	G	12.0	250			43		C	8	N	Y	F	600	1400	5.6	22.0	1	N				9.7		67
98B	3479	S	B	T	99	G	13.4	250			43		C	8	N	Y	F	600	1400	5.6	22.0	1	N				10.7		67
98C	3479	S	B	T	99	G	17.0	250			43		C	8	N	Y	F	600	1400	5.5	22.0	1	N				13.2		67
98D	3479	S	B	T	99	G	18.3	250			43		C	8	N	Y	F	600	1400	5.5	22.0	1	N				14.2		67
98E	3479	S	B	T	99	G	15.5	250			43		C	8	N	Y	F	600	1400	5.5	22.0	1	N				12.2		67
98F	3479	S	B	T	99	G	7.4	250			43		C	8	N	Y	F	600	1400	3.6	18.0	1	N				6.5		67
98G	3479	S	B	T	99	G	6.0	250			43		C	8	N	Y	F	600	1400	3.5	18.0	1	N				5.5		67
98H	3479	S	B	T	99	G	5.3	250			43		C	8	N	Y	F	600	1400	3.3	17.0	1	N				5.0		67
98I	3479	S	B	T	99	G	3.7	250			43		C	8	N	Y	F	600	1400	3.3	17.0	1	N				3.9		67
98J	3479	S	B	T	99	G	5.0	250			43		C	8	N	Y	F	600	1400	3.3	17.0	1	N				4.8		67
98K	3479	S	B	T	99	G	3.2	250			43		C	8	N	Y	F	600	1400	3.1	17.0	1	N				3.5		67
98L	3479	S	B	T	99	G	7.5	250			43		C	8	N	Y	F	600	1400	3.6	18.0	1	N				6.5		67
99	3479	X	M	T		G	30.0	150					B	8	N	N	F	150	1100	7.5	26.0		N	31.0	25.0	65.0			69
100A	3479	N	B	C		G	9.5	350					C	24		N		850	1350				N	5.9		27.5		CATALYST LIFE	
100B	3479	N	B	C		G	9.5	350					C	24		N		850	1350				N	5.9		27.5		CATALYST LIFE	
100C	3479	N	B	C		G	9.5	350					C	24		N		850	1350				N	5.9		27.5		CATALYST LIFE	
100D	3479	N	B	C		G	9.5	350					C	24		N		850	1350				N	5.9		27.5		CATALYST LIFE	
100E	3479	N	B	C		G	9.5	350					C	24		N		850	1350				N	5.9		27.5		CATALYST LIFE	
100F	3479	N	B	C		G	9.5	350					C	24		N		850	1350				N	5.9		27.5		CATALYST LIFE	
100G	3479	N	B	C		G	9.5	350					C	24		N		850	1350				N	5.9		27.5		CATALYST LIFE	
100H	3479	N	B	C		G	9.5	350					C	24		N		850	1350				N	5.9		27.5		CATALYST LIFE	
100I	3479	N	B	C		G	9.5	350					C	24		N		850	1350				N	5.9		27.5		CATALYST LIFE	
100J	3479	X	B	C		G	9.5	350					C	24		N		600	1100							27.5		NEW TYPE CAT	
100K	3479	S	B	T		G	8.0	350					C	24		Y		800	1350							60.0			
100L	3479	S	B	T		G	8.0	350					C	24		Y		600	1350							60.0			
100M	3479	X	B	T		G	10.0	350					C	24		N			1350							40.0			
21	3391	S	D	T		G	5.0	450		150			B	24	N	N	F	500	1650	3.0	24.0		N	4.0	20.0	50.0	18.0	CONTRLS	71
84	3391	N	D	T		G	66.0	425					C	24	N	N	I	425	2400	9.0	27.0		N	16.0		80.0		JUST ON STREAM	71
42A	3323	S	I	T	98	G	11.0	1500					B	4	N	N	I	1500	2000	6.0	38.0		N	2.5				SPK PLUG UV PRB	70
42B	3323	S	I	T	98	G	11.0	1500		6250			B	4	N	N	I	1500	2000	8.0	38.0		N	2.5			1.5	SPK PLUG UV PRB	70
39A	3296	S	D	T	99	G	14.0	240				11.0	C	24		N	F	240	1000	5.0	30.0		Y	9.0	39.0	60.0	8.5	CONTROLS FAN	
39B	3296	S	D	T	99	G	14.0	240					C	24		N	F	240	1000	5.0	35.0		Y	9.0	39.0	60.0	8.5	CONTROLS FAN	
39C	3296	S	D	T	99	G	14.0	240					C	24		N	F	240	1000	5.0	35.0		Y	9.0	39.0	60.0	8.5	CONTROLS FAN	
16	3292	S	M	T	99	G		500					B	8	N	N		500	1500	1.2			N	.3	1.0	2.0		GAS AVAIL	69
33	3079	S	A	T		G	7.6	300	21	375			B	24	N	Y	F			6.8	9.0		N		30.0	60.0	15.0		70
38	3079	S	B	T	99	O	16.0	275	19	500	13500		B	12	N	N	F	275	1200	5.0	22.0		N	18.0	100.0	180.0			67
55A	3079	S	M	T	99	G	3.0	250					C	24	N	Y	F	900	1400	6.1	9.0		N	3.3	23.8	28.8	3.0		69
55B	3079	S	M	T	99	G	4.0	275					C	24	N	Y	F	900	1400	7.3	9.0		N	4.9	31.9	38.5	4.0		69
55C	3079	S	B	T	96	G	11.0	350					C	24	N	Y	F	50	1375	4.4	16.0		N	10.0	25.0	50.0		OVERHEATING	67
15	3069	S	B	T	99	O	7.0	300			1055		C	24	N	Y	I	900	1300	5.0	10.0	89	N	4.2	31.0	65.0	12.0	FAN BURNER	65
64	3000	S	M	T	99	O	6.6	250			1309		C	24	N	Y	I	690	1100			207	N	3.5	35.0	60.0	18.0		69

PROCESS CATEGORY FROM US GOVERNMENT STANDARD INDUSTRIAL CLASSIFICATION
FUME A=ODOR B=HC REDUCTION C=TOXIC FUMES D= SMOKE/PARTICULATES E=OTHER F=A+B G=B+D H=A+D I=A+B+D
FUEL G=NATURAL GAS O=OIL M=EITHER O OR G Z=OTHER PPM = PARTS PER MILLION C1
ORGANICS OUT EITHER PARTS PER MILLION OR R90=90 PERCENT REDUCTION THRU AFTERBURNER
PERFORMANCE S=SATISFACTORY N=NOT SATISFACTORY X=NOT YET RUN
DIMENSIONS IN FEET TEMPERATURES DEGREES F IND/FORCED=INDUCED OR FORCED DRAFT N=NATURAL DRAFT
PRETREAT N=NONE C=CYCLONE D=DILUTION S=SCRUBBER R=ROTOCLONE E=PRECIPITATOR PROBLEMS OS=OUT OF SERVICE

Table 11-1. (Contd-3.)

							WASTE STREAM			CONTAMINANT			AFTERBURNER											FUEL	COST M$				
NO	PROCESS CATEGORY	PERFORMANCE	FUME	THERM/CAT	PERCENT AVAILABLE	FUEL	MSCFM	TEMP	PERCENT OXYGEN	LBS PER HOUR	PPM	BTU PER SCF	BATCH/CONT	HOURS/DAY	PRETREAT	HEAT EXCH	IND/FORCED	TEMP IN	TEMP OUT	EQ DIA	LENGTH	ORG OUT	RECYCLE	MM BTU HOUR	PURCH	INSTD	OP PER YEAR	PROBLEMS	YR
19	2911	S	O	T	95	G	2.0	120	5		8000		C	6	N	N		120	1000	3.5	12.0		N		9.0	19.0	3.0	ENTRAINED LIQ	68
23	2911	S	I	T	99	G	5.0	700					C	10	N	N	I								17.0	20.0	6.0		68
40	2911	N		C		G													800									COST FAILS OS	69
47	2911	S	I	T	99	O		140					B	24	S	Y	I	160	1300	11.0	10.0		N	7.5			26.0		65
48	2911	S	I	T	99	F	7.8	250	7		25000	19.0	C	24	S	N	F	250	1300	5.0	9.0		N	2.1	21.0	28.0		MANY MODS REQUD	69
51	2911	N	I	T	75	O	3.2	470	10	120			B	10	S	Y	F	800	1375	3.5	16.3		N	3.6	27.5	50.0		MANY MECH FAILS	69
52	2911	S	I	T	99	Z	5.0	120	13			40.0	C	24	S	N	N	120	2000	6.0	20.0		N	.8		11.7	.5		69
65B	2911	S	B	T	99	G	1.0	800	15		100		C	24	N	N	F	800	1400	1.8	24.3	R90	N		1.0	10.0		FLAME OUT	67
65C	2911	S	D	T	98	G	3.3	250	10	758	96000	50.0	B	24	S	N	I	250	1800	9.0	8.5	R90	N	12.0	19.8	70.0		FLAME OUT.TEMP	69
66A	2911	N	E	T	97	Z	100.0	1150	1	9999	23000	11.3	C	24	N	Y	F	1150	2000	20.0	31.0		N	225.0				UNBURNED PART.	55
66B	2911	S	E	T	99	M	113.0	525	1		67000	21.6	C	24	E	Y	F	525	1575	30.6	26.0		N	225.0					65
67	2911	N	E	T	95	Z	9.3	315		159	14200		C	24	N	N	F	315	1250	7.0			N	27.0	28.0		67.0		70
68	2911	S	B	T	99	Z							B	*	N	Y	F			6.2	12.0		N						70
102	2911	S	A	T	99	Z	3.0	450				890.0			S		F	450		8.0	10.0		N						58
103A	2911	S	E	T	95	Z	70.0	920			82600	27.0	C	24	C	Y	F	920	1850	19.0	30.0	150	N		330.0			EXCHGR FOULING	66
103B	2911	S	E	T	95	Z	70.0	920			82600	27.0	C	24	C	Y	F	920	1850	19.0	30.0	150	N		330.0			EXCHGR FOULING	66
103C	2911	S	E	T	95	Z	70.0	920			82600	27.0	C	24	C	Y	F	920	1850	19.0	30.0	150	N		330.0			EXCHGR FOULING	66
109	2911	S	E	T	97	G	85.0	500	1		67000	20.0	C	24	E	Y	F	500	1900	20.1	26.5		N	195.0	890.0			EXCHGR FOULING	59
25	2900	S	A	T	99	G							B	24	N	N	F	700	1400	4.0	10.0							CONTROLS	59
27A	2900	S	B	T	99	G	22.0	700	7				C	24	N	N		700	1600	8.0	20.0		N	9.5		150.0	10.0		69
27B	2900	S	B	T	99	G	22.0	700	7				C	24	N	N		700	1600	8.0	20.0		N	9.5		150.0	10.0		6
27C	2900	S	B	T	99	G	22.0	700	7				C	24	N	N		700	1600	8.0	20.0		N	9.5		150.0	10.0		69
27D	2900	S	B	T	99	G	22.0	700	7				C	24	N	N		700	1600	8.0	20.0		N	9.5		150.0	10.0		69
9	2851	S	F	T	99	G	.1	190	1		10000		B		D	N	I	70	1400	2.3	5.0		N	1.9	11.0	20.0	1.0		70
12	2851	S	B	T	99	G	2.0						B	19	S	Y	I	70	1500	2.0	5.0		N	3.0	30.0	100.0	3.5		70
24	2851	S	I	T	99	G											I	500	1100				N					ELECTRICAL	69
34	2851	S	B	C	95	G	1.5						B		S	Y	F	700					N		11.0	21.0	6.0	FIRE CAT POISON	60
37B	2851		B	T		G	9.0								N		I		1200				N			35.0			71
43	2851	S	B	T	98	G	.6	80	20	8	450		B	24	D	N	I	80	1300	2.9	9.5	R85	N	1.9	9.0	40.0	6.0	FL ARRESTOR CNT	67
44	2851	S	A	T	99	G	.1	75	15				B	24	S	N	I	75	1700	2.0	8.0		N	1.1	9.0	30.0	2.9		67
94	2851	S	D	T	99	G	.1	480		55			C	24	N	N	F	480	1550	2.0	7.6		N	.6	18.0	25.0	3.0	INSTRUMENTS	68
58A	2834	S	A	T	95	G	25.0	100	19		40		C	24	S	Y	I	400	1100	7.0	14.0		N	10.0	110.0	363.0	70.0	ROTARY EX.MECH.	66
58B	2834	S	A	T	99	M	25.0	85	20		40		C	24	S	Y	F	850	1150	5.5	20.0		N	11.3	146.0	304.0			71
96	2834	S	A	T		G	1.0	250	16	18			C	24	N	N	I	250	1200	4.3	8.8		N	1.5	15.9	35.0	12.0	CONTROLS	71
2	2815	S	B	T	99	M	31.0	190	17	1050	5800	2.6	C	24	N	Y	F	1060	1300	7.3	29.0	R90	N	9.9			61.0	LEAKS IN DUCTS	67
3	2815	S	F	T	97	Z	33.0	190	19	1150	6300	2.5	C	24	C	Y	F	1090	1250	6.3	22.0	R96	N	10.3	116.0	300.0	97.0	PROCESS UPSET	69
65A	2815	S	B	T	90	G	47.0	[illegible]			4800		C	24	N	Y	F	1000	1300	11.3	40.5	R90	N	12.3	190.0	300.0	35.0	MECH EXCH.	68
30	2800	N	D	T	90	G	20.0	270					C	24	S	N	F	270	850	6.0	16.0		N	.3		35.0	27.0	CORROSION	
49	2800	N	N	C	99		22.0	80	3		7300			24	N	Y	F	960	1100	3.0	8.0	1	N				4.0	COLOR NOX	68
61	2800	S	C	T	97	G							B	24	N	N	F	100	1500	6.0	10.0		N		8.5	10.0		BURNER PLUG OCC	69
79	2800	S	E	C	99		6.7	110	9	193	3000		C	24	S	Y	F	950	1300				N		11.5		7.5		66
78	2800	S	B	T	99	G	9.9	70	14	170		7.5	C	17	S	N	F	70	1800	5.5	25.0	2	N	9.9	59.0	295.0	15.0		71
10A	2700	N	H	I		G	3.0	400					B	20	N	N	F	400	1150	9.0	5.0		N	2.5		10.0		TC FAILURE	62
10B	2700	N	H	T		G	3.0	400					B	20	N	N	F	400	1150	9.0	5.0		N	2.5		10.0		TC FAILURE	62
13	2426	S	O	T		O	6.0	300					C	24	N	Y	F	600	1200	3.8	11.0		N	3.9	62.0	85.0	19.0	REFRACTORY	69
17A	2095	N	G	T	99	G	3.7	350	14	140			C	8	R	N	F	350	900	4.5	8.0		Y	3.0		21.0	1.0	HC REDUCTION	60
17B	2095	N	G	T	99	G	8.0	400	16	800			C	16	N	Y	F	400	950	8.0	15.0		N	5.0	20.0	39.0	12.0	HC REDUCTION	67
17C	2095	X	G	T		G	12.0	400	16	800			C		N	Y	I	1200	1400		19.0				75.0	150.0			71
20	2095	S	D	T	97	G	5.0	300	12				B	8	S	N	F	250	1200	4.5	10.0		N	6.0	5.0	10.0		FAN CONTRLS	70
70A	2095	S	I	T	99	G	2.0	135	21			.7	C	8	C	N	F	135	1200	1.9	7.5		N	.7	3.1	4.0	.8		70
70B	2095	S	I	T	99	G	2.0	135	21			.7	C	8	C	N	F	135	1200	1.9	7.5		N	.7	3.1	4.0	.8		70

PROCESS CATEGORY FROM US GOVERNMENT STANDARD INDUSTRIAL CLASSIFICATION
FUME A=ODOR B=HC REDUCTION C=TOXIC FUMES D= SMOKE/PARTICULATES E=OTHER F=A+B G=B+D H=A+D I=A+B+D
FUEL G=NATURAL GAS O=OIL M=EITHER O OR G Z=OTHER PPM = PARTS PER MILLION CI
ORGANICS OUT EITHER PARTS PER MILLION OR R90=90 PERCENT REDUCTION THRU AFTERBURNER
PERFORMANCE S=SATISFACTORY N=NOT SATISFACTORY X=NOT YET RUN
DIMENSIONS IN FEET TEMPERATURES DEGREES F IND/FORCED=INDUCED OR FORCED DRAFT N=NATURAL DRAFT
PRETREAT N=NONE C=CYCLONE D=DILUTION S=SCRUBBER R=ROTOCLONE E=PRECIPITATOR PROBLEMS OS=OUT OF SERVICE

Table 11-1. (Contd-4).

NO	PROCESS	CATEGORY	PERFORMANCE	FUME	THERM/CAT	PERCENT AVAILABLE	FUEL	********WASTE STREAM******** MSCFM	TEMP	PERCENT OXYGEN	CONTAMINNT LBS PER HOUR	CONTAMINNT PPM	BTU PER SCF	BATCH/CONT	HOURS/DAY	PRETREAT	HEAT EXCH	IND/FORCED	******AFTERBURNER****** TEMP IN	TEMP OUT	EQ DIA	LENGTH	ORG OUT	RECYCLE	FUEL MM BTU HOUR	****COST M$ PURCH	INSTD	OP PER YEAR	****PROBLEMS****	YR
70C	2095		S	I	T	99	G	2.0	135	21			.7	C	8	C	N	F	135	1200	1.4	7.5		N	.7	3.1	4.0	.8		70
70D	2095		S	I	T	99	G	2.0	135	21			.7	C	8	C	N	F	135	1200	1.4	7.5		N	.7	3.1	4.0	.8		70
70E	2095		S	I	T	99	G	2.0	135	21			.7	C	8	C	N	F	135	1200	1.4	7.5		N	.7	3.1	4.0	.8		70
70F	2095		S	I	T	99	G	2.0	135	21			.7	C	8	C	N	F	135	1200	1.4	7.5		N	.7	3.1	4.0	.8		70
73A	2095		S	B	T	99	G	1.5	350	15	3	250		C	16	N	N	F	350	1500	4.0	20.0		N	1.5	25.0	50.0	4.0		70
73B	2095		S	B	T	99	G	.6	350	16	1	250		C	16	N	N	F	350	1000	2.0	4.0		N	.1	4.0	8.0	.5	STICKING DAMPER	62
81A	2095		S	G	T		G	1.6	220		1	1400		B	8	N	N	F		1200	1.6	5.8		N		3.5	7.2	.5		71
81B	2095		S	G	T		G	1.6	220		1	1400		B	8	N	N	F		1200	1.6	5.8		N		3.5	7.2	.5		71
81C	2095		S	G	T		G	1.6	220		1	1400		B	8	N	N	F		1200	1.6	5.8		N		3.5	7.2	.5		71
81D	2095		S	G	T		G	1.6	220		1	1400		B	8	N	N	F		1200	1.6	5.8		N		3.5	7.2	.5		71
81E	2095		S	G	T		G	1.6	220		1	1400		B	8	N	N	F		1200	1.6	5.8		N		3.5	7.2	.5		71
81F	2095		S	G	T		G	1.6	220		1	1400		B	8	N	N	F		1200	1.6	5.8		N		3.5	7.2	.5		71
95A	2095		S	B	T	99	G	3.7	475	11		393		C	16	C	N	F	475	1300	2.6	12.0		N	3.0			4.0	FLAME ROD	69
95B	2095		S	B	T	99	G	3.7	475	11		393		C	16	C	N	F	475	1300	2.6	12.0		N	3.0			4.0	FLAME ROD	69
101A	2095		S	F	T	99	G	.6	450					B	8	C	N	F	450	1250	1.3	7.0		N	.7	10.0	16.0	5.3		65
101B	2095		S	F	T	99	G	.6	450					B	8	C	N	F	450	1250	1.3	7.0		N	.7	10.0	16.0	5.3		65
101C	2095		S	F	T	99	G	.6	450					B	8	C	N	F	450	1250	1.3	7.0		N	.7	10.0	16.0	5.3		65
101D	2095		S	F	T	99	G	.6	450					B	8	C	N	F	450	1250	1.3	7.0		N	.7	10.0	16.0	5.3		65
28	2094		S	B	T		G	30.0	100				3.5	B	20	N	Y	F	550	1200					15.0	99.0	193.0	60.0		69
29	2094		S	B	T	95	O	20.0	150					C	19	S	N	F	150	1250	12.0	20.0		Y	36.0	25.0	40.0	30.0	SLOW HEAT-UP	69
53	2094		S	A	T	99	G													1300	5.5	6.0								65
7A	2013		S	D	T	99	G	2.3	190	19				B		N	N		190	800	2.0	8.0		N		8.5	12.0	7.0	REFRACTORY	69
7B	2013		S	D	T	99	G	2.3	190	19				B		N	N		190	800	2.0	8.0		N		8.5	12.0	7.0	REFRACTORY	69
7C	2013		S	D	T	99	G	2.3	190	19				B		N	N		190	800	2.0	8.0		N		8.5	12.0	7.0	REFRACTORY	69
19A	2013		S	D	T	99	G	2.3	200					B	18	N	N	F		1200	2.5	6.0		N	.5	5.0	8.0			
19B	2013		S	D	T	99	G	2.3	200					B	18	N	N	F		1200	2.6	6.0		N	.5	5.0	8.0			
19C	2013		S	D	T	99	G	2.3	200					B	18	N	N	F		1200	2.5	6.0		N	.5	5.0	8.0			
19D	2013		S	D	T	99	G	2.3	200					B	18	N	N	F		1200	2.5	6.0		N	.5	5.0	8.0			
19E	2013		S	D	T	99	G	2.3	200					B	18	N	N	F		1200	2.5	6.0		N	.5	5.0	8.0			
19F	2013		S	D	T	99	G	2.3	200					B	18	N	N	F		1200	2.5	6.0		N	.5	5.0	8.0			
31A	2000		S	B	T	98	G		400					B	16	N	N	I	400		3.0	5.0		N						67
31B	2000		S	B	T	98	G		400					B	16	N	N	I	400		3.0	5.0		N						67
31C	2000		S	B	T	98	G		400					B	16	N	N	I	400		3.0	5.0		N						67
46	2000		N	D	T	99	G	3.0						B	6	N	N	F			3.3	8.0		N	1.9	3.5	5.0	2.9	OP COST PERFMNC	70

Table 11-2 TOTAL ANNUAL COST

	Catalytic			Thermal		
	Pre-engineered No Exchanger	Custom Unit No Exchanger	Custom Unit 40% Exchanger	No Exchanger	40% Exchanger	60% Exchanger
Purchase, $	22,000	38,000	48,000	22,000	46,000	74,000
Installed, $	22,000	38,000	48,000	22,000	46,000	74,000
Total Capital, $	44,000	76,000	96,000	44,000	92,000	148,000
Depreciation, $	4,400	7,600	9,600	4,400	9,200	14,800
Interest on Capital, $	4,000	6,800	8,600	4,000	8,300	13,300
Catalyst replacement, $	7,500	7,500	7,500			
Maintenance, $	900	1,500	1,900	900	1,800	3,000
Fuel, $	16,000	16,000	7,100	32,300	17,900	10,100
Blower power, $	500	500	800	200	500	800
Total Annual Cost, $	33,300	39,900	35,500	41,800	37,700	42,000

Chapter 12. COMBUSTION PRINCIPLES

In order to logically design an afterburner, one should understand the principles which govern the oxidation process, especially as it is utilized in dilute fume afterburners. Therefore, in the sections which follow we discuss the oxidation process with its associated heat generation and loss; the meaning of ignition temperature and flammability limits; the reasons why some of the waste gas must bypass the flame; the mechanisms involved in the oxidation process; estimated oxidation rates for various pollutants; the role of mixing processes; and the fate of various heteroatoms. It is not our intention to present a complete treatise on combustion or combustion chamber and burner design. Good general references are books by Spaulding,12-1) Bradley,12-2) and Minkoff and Tipper.12-3)

12.1 The Oxidation Process - A Balance Between Heat Generation and Removal

Steady state combustion or thermal oxidation processes are defined by a balance between heat generation by the exothermic reaction and heat removal from the local region where reaction is occurring. Actually, combustion is far from a true steady state operation, as any observation of flickering flames will indicate, but one can view flames and thermal oxidation as overall steady state processes. We shall use this approach to define ignition temperatures, flame temperatures and combustible limits in order to provide an understanding of the basic requirements of afterburner design. The treatment is brief, but an entire book by Vulis12-4) is available on the subject.

Oxidation involves bimolecular reactions between the combustible compound and the oxidizer. As shown in Section 12.2, the oxidizer is usually not molecular oxygen but this is not important for the discussion here. The rate at which oxidation proceeds can be approximately represented by the overall expression

$$r = k[\chi_{O_2}][\chi_{\text{combustible}}] \qquad (12\text{-}1)$$

where r is the rate at which the combustible compound is oxidized, moles/cm^{3}sec; k is the rate constant, moles/cm^{3}sec; and χ is the mole fraction. The rate constant, k, is an exponential function of temperature

$$k \simeq A\, e^{-E/RT} \qquad (12\text{-}2)$$

with activation energy, E, typically in the range of 10-60 kcal/gmole°K for this overall rate constant. This means that oxidation rates are extremely temperature sensitive. For an activation energy of 30 kcal/mole°K, the reaction rate will increase by 103 times for an increase of 50°F at 80°F and by 1.52 times for the same 50°F increase at 1340°F. Equation 12-1 shows that the oxidation rate is also dependent on the concentration of combustible. Since combustible is consumed in the reaction, the rate decreases an oxidation proceeds. This leads to the familiar S curve (Figure 12-1) of extent of reaction (i.e. fraction of the hydrocarbon contaminant, which has been oxidized to CO_2 and H_2O) versus temperature for a constant reactor (combustor) residence time. The curves in Figure 12-1 were computed using the CO oxidation kinetics reported by Yuster, et al,(12-5) and given as Equation 12-37. (The CO oxidation kinetics should be considered suspect at these low temperatures, however.) Oxygen is assumed to be in large excess, so its concentration is constant. (Section 12.3.2 discusses the various rates measured for CO combustion.) As can be seen, for a given residence time and temperature, the plug flow (zero backmixing) reactor gives more complete destruction of combustible than the stirred tank reactor (complete backmixing) in which the concentration is everywhere taken equal to the exit concentration.

The extent of reaction can also be expressed as a dimensionless heat generation (by the exothermic oxidation reaction), if the heat of combustion is taken as constant. The upper section of Figure 12-2 shows the S curve of this dimensionless heat generation versus temperature for a single well stirred reactor with 0.1 seconds residence time. Steady state operation requires that this heat generation be balanced by heat losses and convection. If heat losses can be ignored, the dimensionless convective heat removal rate (as a fraction of the maximum heat generation rate) is

$$Q_{\text{convective heat removal}} = \frac{T - T_{\text{inlet}}}{X_{\text{inlet}} \Delta H / C_p} \qquad (12\text{-}3)$$

where T = temperature within the stirred tank reactor (combustor)
X_{inlet} = mole fraction of combustible in the incoming stream
ΔH = heat of combustion, Btu/lb mole (for complete conversion to CO_2 and H_2O)
C_p = average of combustible heat capacity of flowing stream, Btu/lb mole°F

The straight line given by this equation is shown as the dotted line in the same coordinates as those used for S-shaped heat generation curve. The line shown was calculated for a waste stream inlet (preheat) temperature of 800°F and for a stream hydrocarbon concentration equal to one-fourth the lower explosion limit (1/4 LEL). At steady state, heat generation and removal must be equal. Thus, points A, B and C, the intersections of the heat generation curve with the heat removal line represent possible steady state operating conditions for the system.

Figures 12-1 and 12-2 follow

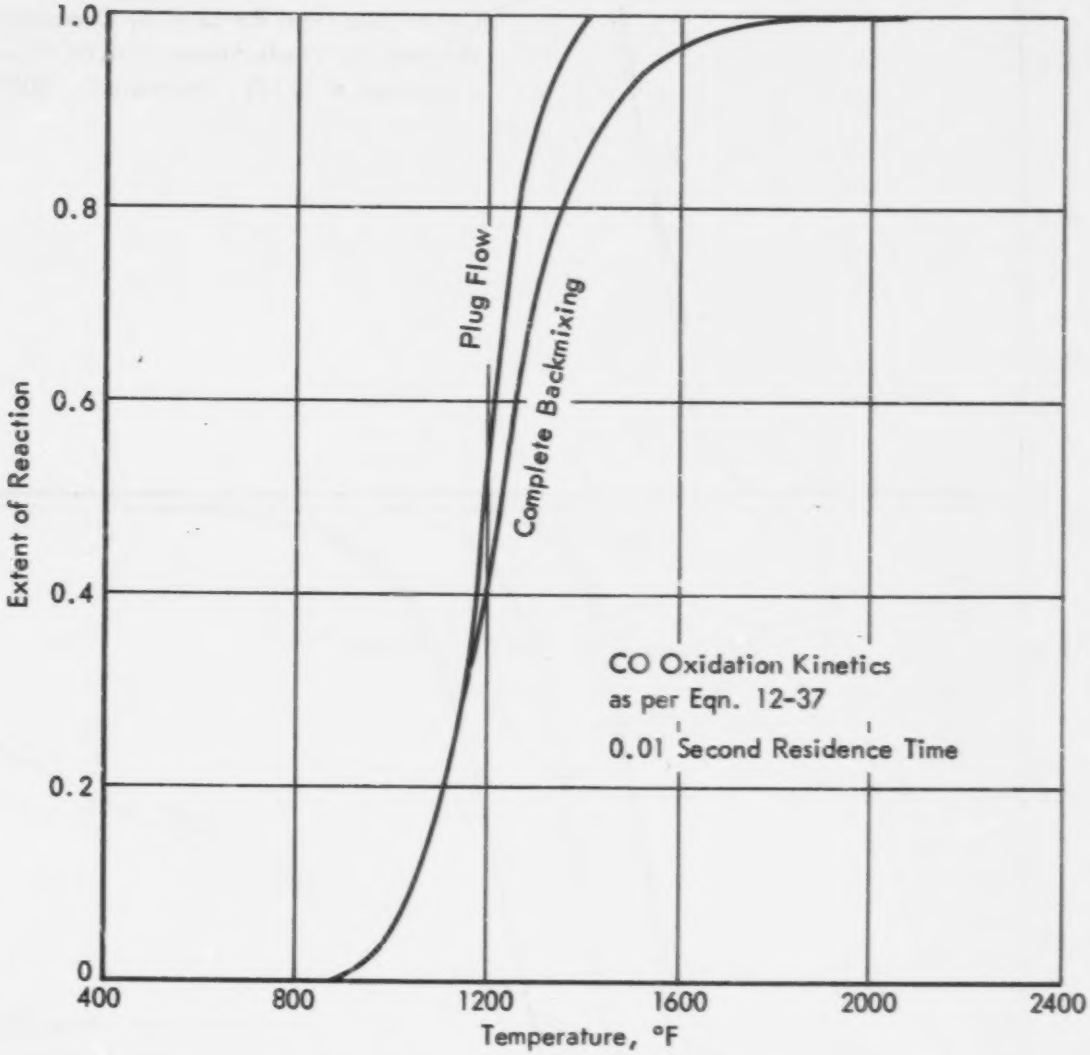

Figure 12-1. EFFECT OF TEMPERATURE AND BACKMIXING ON EXTENT OF REACTION

180b

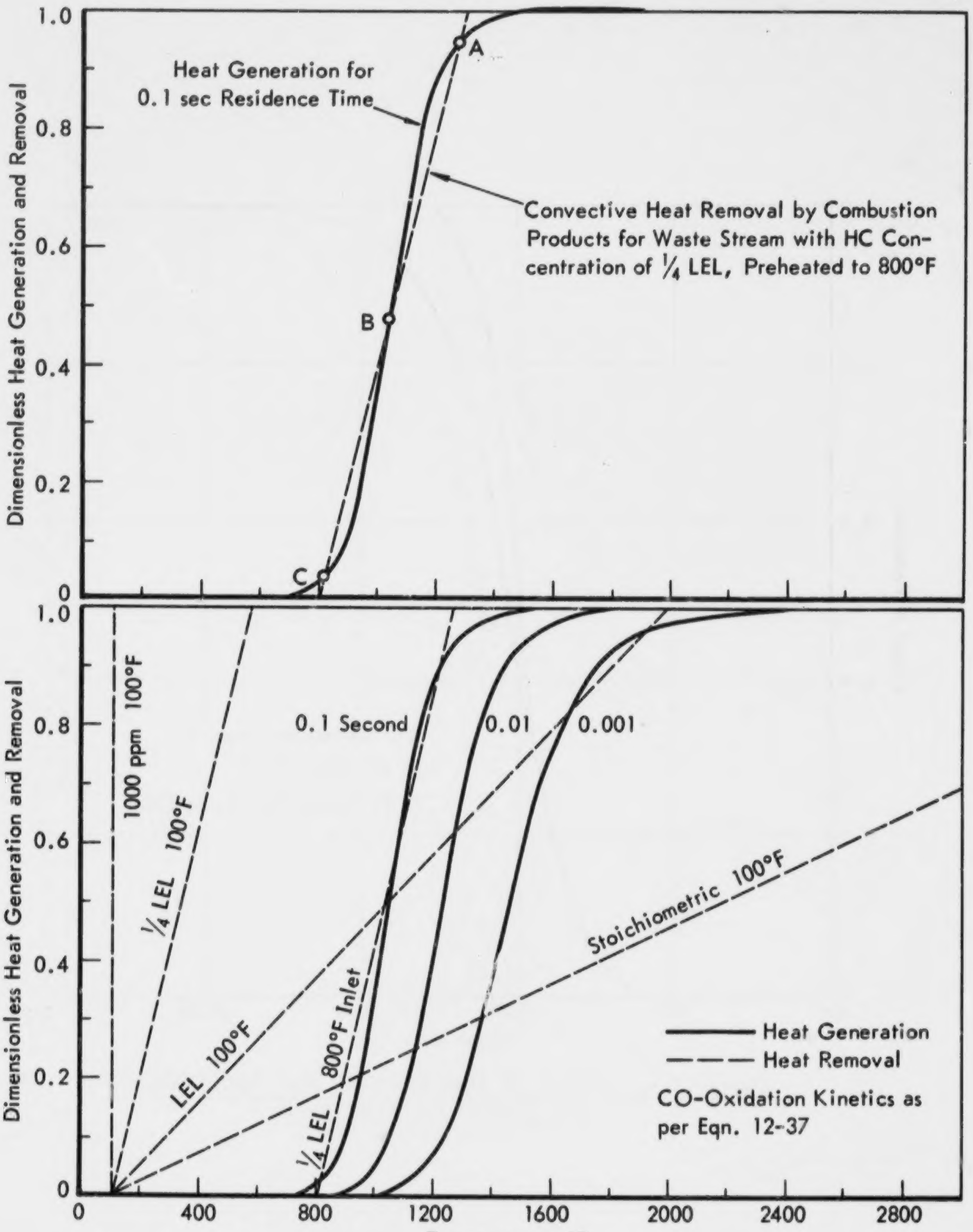

Figure 12-2. HEAT GENERATION AND HEAT REMOVAL

Effects of Temperature and Residence Time on Generation.
Effects of Concentration and Preheat on Removal.

S-14121
67784

180C

The lower portion of Figure 12-2 shows heat generation curves for three residence times, and five heat removal lines, representing four values of hydrocarbon concentration in the waste stream and two preheat temperatures. The LEL for CO is taken as 12.5%v, ΔH is taken as 1.217 x 10^5 Btu/lb mole°F. As can be seen, the slope of the heat removal line increases as the concentration decreases. For very low concentrations a nearly vertical line is obtained. If heat losses occur and these losses are proportional to the combustor temperature, the slope of the heat removal line will be greater for a given concentration.

Each intersection between a line and curve represents a possible steady state condition for the residence time specified on the heat generation curve, and the hydrocarbon concentration and preheat temperature specified on the heat removal line involved in the intersection. With a low combustible concentration and low inlet temperature, the oxidation rate and extent of reaction are near zero. As the inlet temperature is raised, the heat removal line shifts to the right and will eventually intersect the generation line at an arbitrarily high extent of reaction. As discussed below, this defines the ignition temperature.

High concentration heat removal lines have three intersections with the heat generation curve. The central one is unstable and defines the ignition temperature. If the temperature of the mixture is raised above this, the high temperature steady state will be attained.

Figure 12-2 is only presented to give a qualitative picture of the coupling between heat transfer and reaction. The temperatures shown, however, are much lower than found to be necessary in practice (see Section 3.1.2.3). A major reason for this is that heat losses have not been included in Figure 12-2. These are very important, especially with a dilute fume stream incapable of generating much heat itself (near vertical heat removal line). Heat losses (conduction and radiation) limit the extent to which reaction time can be increased to compensate for a low reaction rate. Reactions must occur rapidly or the heat generated will be lost to the surroundings instead of raising the adjoining fuel and air mixture to its ignition temperature. Another problem may be that the reaction rates used to calculate Figure 12-2 were too high. In Section 12.3.2 we compare various rate expressions for CO oxidation.

12.1.1 Ignition Temperature

The term ignition temperature can be understood in light of the discussion in Section 12.1 and Figure 12-2. For high concentrations of combustibles there are three intersections of a heat generation curve with a given heat loss line, and the middle (unstable) operating point will define the ignition temperature. If the mixture is raised above this, the combustor will jump to the upper operating point characterized by nearly complete conversion and a high temperature.

Low combustible concentrations (typical of afterburner applications) give a nearly vertical heat removal line since very little heat is generated by the oxidation reaction. In this case there is only one intersection of the generation and loss lines and ignition temperature is more difficult to define. One way is to associate it with the temperature required for 90% (or some other value) conversion. Another is to take the point of inflection (maximum slope) on the heat generation curve (~50% conversion). For high activation energy reactions these are not far different.

But it should be apparent that these "ignition temperatures" will be extremely sensitive to experimental conditions. The measured value would be affected by:

Changes in heat removal line

1. The concentration of combustible component (and oxidant)
2. The inlet temperature
3. The rate of heat loss from the apparatus

Changes in the heat generation (extent of reaction) line

1. The combustor residence time and flow pattern (deviations from complete backmixing)

2. Inhibition or catalysis by vessel walls so results depend on vessel geometry and material of construction

Therefore, published values of autoignition temperatures should be used only for general guidance as indicated by the wide range of temperatures for many compounds.

12.1.2 Combustion Limits and Flame Temperatures

The upper and lower flammability or explosive limits (maximum and minimum concentrations of fuel in air which will support combustion) are also determined by a balance between heat generation and loss. If combustion is to be self-sustaining, heat generation by reaction in an ignited region must exceed heat losses by enough to raise neighboring unburned fuel and air mixture above their "ignition temperature." Empirically, it has been observed[12-8]) that sustained combustion requires the release of at least 10 kcal per mole of combustion products above that needed to raise these products to the ignition temperature. This provides the driving force for transport of heat to unburned regions.

The rate of combustion and flame temperature are greatest for a stoichiometric mixture of oxygen and fuel since no diluent (excess oxygen or fuel) is present to absorb heat. Also, reactant concentrations for the bimolecular reactions are optimal. For pure hydrocarbons, this requires (x + 1/4y) moles of oxygen per mole of hydrocarbon C_xH_y. When air is used, the 79/21 moles of nitrogen per mole of oxygen serve to lower both rate and flame temperature. For most hydrocarbons, the adiabatic flame temperature for a stoichiometric mixture of air and hydrocarbon is ~4000°F. As heat sink (excess air or fuel) is added, both the flame temperature and oxidation rate drop. For most hydrocarbons in air, combustible mixtures at room temperature range from 1/2 → 3 times the stoichiometric concentration as measured in mole percent. Table 12-1 gives a few representative values for the reported lower explosive limit LEL, heat content, and temperature rise for the minimum explosive mixture. As can be seen, the LEL for these compounds vary, but the corresponding heat content and ΔT are not very different. This would be expected from the discussion above.

Table 12-1. HEAT CONTENT OF VARIOUS FUME STREAMS AT THE LOWER EXPLOSIVE LIMIT

Bases: 379 SCF/lbmole
C_p = 8.5 Btu/lbmole°F

Combustible	LEL at Room Temperature	Heat Content	ΔT Corresponding to Combustion
Methane	5.0%v	45.7 Btu/SCF	2040°F
Propane	2.1%v	48.7 Btu/SCF	2190°F
Hexane	1.2%	52.9 Btu/SCF	2380°F
Toluene	1.2%	51.3 Btu/SCF	2310°F
Methyl ethyl ketone	1.9%	49.4 Btu/SCF	2220°F

Table 12-1 also shows that stable flame temperatures for these representative combustibles cannot be less than ~2100°F → 2400°F. Imperfect mixing and heat losses result in slightly more fuel than the LEL being required to sustain combustion. If air dilution is increased to give lower flame temperatures (mixtures below the LEL), the flame will be extinguished.

This then is the reason why unburned fuel cannot be completely mixed with a dilute fume stream in order to achieve a 1400°F-1500°F operating temperature. The heat content of the fuel will not be released since the maximum heat generation is not sufficient to insure ignition of the cold mixture. For the same reason, a stream containing ~50-60% of the LEL cannot be destroyed without burning supplemental fuel despite the fact that the heat content of the fume is sufficient to raise it to 1400-1500°F. The incoming fume must be raised to its "ignition temperature" in order to release the available heat and oxidize the pollutants. Preheating the fume will minimize the required supplemental fuel.

12.1.3 Dilute Fume Must Bypass Flame

The discussion in the above sections shows why dilute fumes must be mixed with the combustion products from burning the supplemental fuel rather than with the unburned fuel itself. Where possible, oxygen present in the fume should be used for combustion of the fuel in order to avoid introducing outside air which requires additional fuel to raise it to the operating temperature (see Section 3.1.3.1). However, the remaining fume must be mixed in downstream of the combustion zone. This amounts to ~50% of the total fume in the case of 80°F fume with negligible heating value, 100% excess air in the burner (near the LEL), and 1400°F combustion chamber temperature. More fume must bypass if the burner is to operate with less excess air or if less fuel is needed because the fume is preheated or has a significant heating value. If the fume enters at 400°F rather than 80°F, ~65% of the fume must bypass the burner, other conditions remaining the same. Schematically, the process can be viewed (Figure 12-3) as three distinct steps.

1. Supplemental fuel combustion (generates heat)
2. Mixing of fume and combustion products
3. Oxidation of the combustible pollutants in the fume (hold fume at temperature for required time)

The way these three steps are accomplished in available afterburners varies considerably and leads to a large range of effectiveness from complete conversion to CO_2 and H_2O to increasing rather than decreasing the amount of pollutants in the fume stream. This is discussed in more detail in Chapters 13, 14, and 15.

If the fume stream contains a high enough concentration of combustibles to sustain combustion at the temperature it is supplied, it can be treated as a premixed (or raw) fuel stream. In this case only a single combustion step is required since the fume itself supplies all the required heat. For safety reasons, such high concentration streams are likely to have little, if any, oxygen so combustion air must be provided.

Figure 12-3 follows

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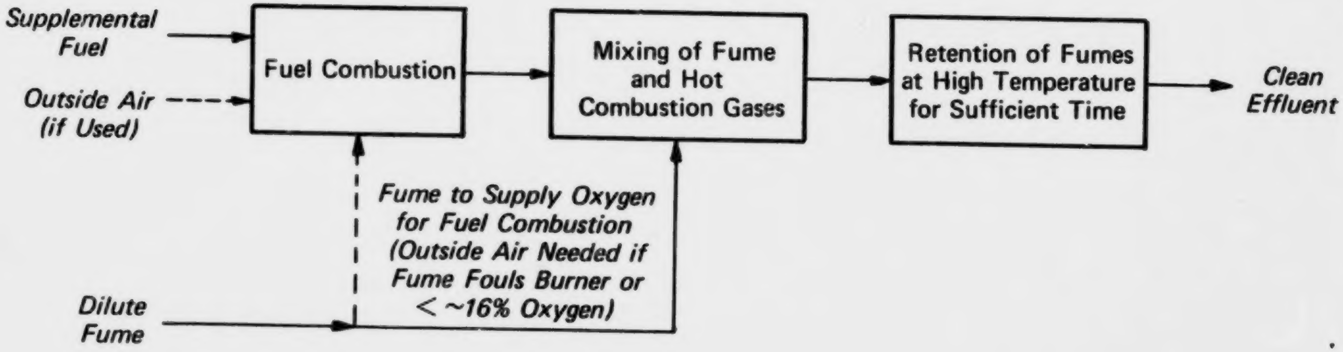
Supplemental Fuel
Outside Air (if Used)
Fuel Combustion
Mixing of Fume and Hot Combustion Gases
Retention of Fumes at High Temperature for Sufficient Time
Clean Effluent
Fume to Supply Oxygen for Fuel Combustion (Outside Air Needed if Fume Fouls Burner or < ~16% Oxygen)
Dilute Fume

Figure 12-3. STEPS REQUIRED FOR SUCCESSFUL INCINERATION OF DILUTE FUMES

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12.2 Mechanisms of Oxidation and Combustion

Dilute fume incineration involves oxidation reactions at combustible concentrations and temperatures much lower than are normally studied in combustion research. As discussed above, flames cannot be supported and kinetic rates and mechanisms determined using flames may not be applicable. In addition, the normal operating temperatures 1200-1500°F are of the same order as the transition between "cool flame" kinetics (where peroxides are important chain carriers) to hot flame kinetics (involving excess concentrations of O, OH, and H). High temperature mechanisms are discussed first.

12.2.1 Free Radical Flame Reactions

Within flames, oxidation does not take place by reaction of molecular oxygen with fuel but rather through free radical mechanisms. This reflects the near zero activation energy of free radical oxidation reactions. For instance, the activation energy for the oxidation of carbon monoxide by OH radicals

$$CO + OH \rightarrow CO_2 + H \tag{12-4}$$

is only 1 kcal/mole, whereas that for the direct oxidation by molecular oxygen

$$CO + O_2 \rightarrow CO_2 + O \tag{12-5}$$

is 48 kcal/mole. In addition, in flames, one finds OH, O, and H radical concentrations orders of magnitude higher than would be predicted by the dissociation equilibrium for H_2O, O_2, and H_2. This can be explained by the high rates for the chain branching reactions (two radicals produced for every one involved).

$$H + O_2 \rightarrow OH + O \tag{12-6}$$

$$O + H_2O \rightarrow OH + OH \tag{12-7}$$

$$O + H_2 \rightarrow OH + H \tag{12-8}$$

$$H + H_2O \rightarrow H_2 + OH \tag{12-9}$$

A partial equilibrium is quickly established between radicals via these reactions and the reverse of these. However, the much slower three-body recombination reactions

$$OH + O + M \rightarrow H_2O + M \tag{12-10}$$

$$H + H + M \rightarrow H_2 + M \tag{12-11}$$

$$O + O + M \rightarrow O_2 + M \qquad (12\text{-}12)$$

where M represents the third (energy absorbing) body

do not return the system to complete equilibrium with molecular species until sometime after all fuel has been consumed. Typically, peak OH concentrations in flames (~3000°F) are in the range .01 → .05 mole fraction and may be as high as .10. The corresponding equilibrium mole fractions are one or two orders of magnitude lower, ~.001 → .002. At lower temperature, the equilibrium values are even lower, e.g., OH mole fraction $\sim 10^{-7}$ @1400°F. Combustion reactions can be quenched by a drop in temperature which drastically reduces rates or by the presence of a high concentration of third bodies, e.g., high surface area or many inert molecules.

Radicals can be generated by initiation reactions involving slow reactions between fume and fuel. However, in continuous flow combustors these initiation steps are not very important since radicals are carried from burning regions to unburned regions by mixing and diffusion. (Flame stabilizing bars or discs are used to form stable vortices in their wake which provide backmixing of heat and radicals. This stabilization can also be accomplished aerodynamically. Chapter 13 contains additional discussion of flame holding.)

Initial attack of the vaporized fuel molecule in an oxidizing flame is probably by an OH radical. For methane, this is

$$CH_4 + OH \rightarrow H_2O + CH_3 \qquad (12\text{-}13)$$

Then the chain continues

$$CH_3 + O_2 \rightarrow CH_2O + OH \qquad (12\text{-}14)$$

or

$$O + CH_3 \rightarrow CH_2O + H \qquad (12\text{-}15)$$

forming formaldehyde. This reacts further to give CO

$$CH_2O + OH \rightarrow H_2O + CHO \qquad (12\text{-}16)$$

and

$$CHO + OH \rightarrow H_2O + CO \qquad (12\text{-}17)$$

which is oxidized to CO_2 via reaction 12-4

$$CO + OH \rightarrow CO_2 + H \qquad (12\text{-}4)$$

All evidence points to CO oxidation as the slow step in oxidation of hydrocarbons. As discussed in Section 3.1.2.3, this rate often controls the required residence time in afterburners.

Hydrocarbons composed of more than one carbon will undergo a similar sequence, with organic acids and aldehydes appearing as intermediates. With hydrocarbons of more than 3 or 4 carbons, the radical formed in the equivalent of reaction 12-13 is unstable and is likely to undergo pyrolytic decomposition (endothermic) and/or dehydrogenation.

$$RCH_2CH_2\cdot \rightarrow R'CH = CH_2 + R''\cdot \qquad (12\text{-}18)$$

or

$$RCH_2CH_2\cdot \rightarrow RCH = CH_2 + H \qquad (12\text{-}19)$$

As a result, aldehydes containing more than two carbon atoms are not likely to be found in flames. Aldehyde concentration in flames is generally very low since all steps except the oxidation of CO to CO_2 are very fast at flame temperatures. If concentrations of O and OH are low (poor mixing or fuel rich flames), dehydrogenation may be accompanied by polymerization to yield material of a high C/H ratio--smoke and soot. Steam is often injected to reduce soot formation. This helps to increase OH concentration via reactions 12-7 and 12-9.

12.2.2 Low Temperature Oxidation

At very low temperatures oxidation rates can be taken as zero. However, for temperatures between ~500°F and the "ignition temperature" (see Section 12.1.1) relatively slow chain reactions occur which yield "cool flames" for combustible mixtures. As in high temperature flames, OH radicals are a major oxidizing species but in this case their concentration is much lower and peroxide radicals (HO_2) play a major role. Combustion times at 500-700°F are on the order of 10-30 minutes rather than milliseconds as in normal flames.

To illustrate cool flame mechanisms we shall again use methane as an example. Initiation occurs by

$$CH_4 + O_2 \rightarrow CH_3 + HO_2 \qquad (12\text{-}20)$$

which is very slow. A chain reaction is set up with initial attack on CH_4 via

$$CH_4 + OH \rightarrow H_2O + CH_3 \qquad (12\text{-}21)$$

and

$$CH_4 + HO_2 \rightarrow H_2O_2 + CH_3 \qquad (12\text{-}22)$$

Then methyl radicals react with oxygen to give formaldehyde.

$$CH_3 + O_2 \rightarrow CH_2O + OH \qquad (12\text{-}23)$$

which leads to branching by

$$CH_2O + O_2 \rightarrow CHO + HO_2 \qquad (12\text{-}24)$$

Formaldehyde can also react with OH and HO_2 radicals

$$CH_2O + OH \rightarrow H_2O + CHO \qquad (12\text{-}25)$$

$$CH_2O + HO_2 \rightarrow H_2O + CHO \qquad (12\text{-}26)$$

The oxidation continues to CO via

$$CHO + O_2 \rightarrow CO + HO_2 \quad . \qquad (12\text{-}27)$$

CO oxidation is very slow at low temperatures but probably occurs via an ozone intermediate or other excited species.

$$CO + O + M \rightarrow CO_2 + M \qquad (12\text{-}28)$$

As can be seen, some of the reaction steps are the same at low and high temperatures. However, the rates are much slower and concentrations of OH and O are much lower. As a result, the less active peroxide radicals play an important role. Also, aldehydes are much more prevalent and provide for chain branching through reaction 12-24.

Higher hydrocarbons follow a similar sequence with corresponding aldehydes, acids, alcohols, and peroxides formed as intermediates. As with higher temperature processes, large hydrocarbon molecules tend to decompose following the initial hydrogen abstraction step.

12.2.3 Thermal Oxidation in the Absence of Flame

The above discussion shows the important role free radicals play in the oxidation of hydrocarbons to CO_2 and H_2O. It has also been pointed out that radical concentrations in flames are one or two orders of magnitude above equilibrium values with correspondingly high reaction rates.

Dilute fume incineration, however, involves thermal oxidation in the absence of flame. The reasons for this were discussed in Section 12.1.3. In afterburners radical concentrations are expected to be much closer to equilibrium values once excess levels from the supplemental fuel burner have decayed. Chain branching reactions dominate only while fuel is being consumed. Within the combustion chamber combustible concentration in the fume will determine the amount by which radical concentrations can exceed the overall equilibrium values set by the temperature and concentrations of water and oxygen. Radicals generated by reactions such as 12-14 or remaining from the fuel burner will tend to recombine by reactions 12-10 → 12-12 rather than lead to branching. The lower the concentration of combustibles, the nearer the radical concentrations will be to overall equilibrium values since chain terminating collisions of radicals with third bodies are more likely than reacting collisions with hydrocarbon molecules. In Sections 12.3.1 and 12.3.2 CO and methane oxidation rates are compared for flame conditions (excess free radicals) and for radical concentrations at equilibrium values. Experience with afterburner operation indicates that actual oxidation rates are somewhere between these extremes as would be expected from the above discussion. Also, very dilute fumes are more difficult to clean up than more concentrated fumes.

The concept of "flame contact" would appear to have much value since the excess free radicals formed in the supplemental fuel flame could be utilized to oxidize the combustible pollutants in the fume. Typical decay times for these excess radicals are on the order of .01 → .03 seconds. However, this is of the same order as CO oxidation in the flame and attempts to rapidly mix fume and flame on this time scale will lead to flame quenching. As discussed in Chapter 14, normal mixing times are an order of magnitude longer.

12.3 Oxidation Kinetics

Once fume and combustion gases have been completely mixed and brought to a temperature above the ignition temperature, the oxidation reactions must be given sufficient time to occur before the gases are quenched in a heat exchanger, by air dilution, or by exiting through a stack. As is discussed in Section 12.4, fuel combustion in flames is essentially controlled by mixing rates, and oxidation kinetics usually are of secondary importance. In afterburners, however, low temperatures and reduced radical concentrations make oxidation rates comparable to mixing rates. In this section we give some estimates for these required times. Data is extremely sparce, especially at temperatures and concentrations of interest in fume incineration. Therefore, a certain amount of design flexibility and testing is required to allow even the most stringent effluent criteria to be met.

12.3.1 Hydrocarbon Destruction

The most common use of afterburners is to destroy solvents or other hydrocarbon fumes and odors. In the combustion of fuels it has been found that the oxidation of $CO \rightarrow CO_2$ (reaction 12-4 in Section 12.2.1) is the slowest step and the kinetics of the steps leading from the original hydrocarbon to CO can be ignored as a first approximation.$^{12-7)}$ Temperature levels in afterburners are much lower than in flames so all rates are lower, but carbon monoxide oxidation again seems to control. A possible exception is methane oxidation which has a very high activation energy (~60 kcal/mole), and except at flame temperatures (>3000°F), the oxidation of methane to CO is considered rate limiting.$^{12-8)}$

The curves shown in Figures 12-4 and 12-5 represent$^{12-9)}$ experimental data taken by Surface Combustion Division of Midland Ross Corporation. Measurements were made in the absence of a flame and at temperatures and concentrations of interest. Prevaporized solvent was added to air heated to the desired operating temperature. Mixing was designed to be nearly instantaneous, and plug flow was maintained in the reactor. As can be seen, the rate of hydrocarbon disappearance was taken to be first order in solvent concentration and somewhat dependent on solvent type. Toluene, which is often reported to have a very high ignition temperature, disappeared more slowly than cyclohexane or hexane.

Figure 12-5 shows the rate at which toluene is oxidized as a function of afterburner temperature. Rates are slow below 1400°F but increase rapidly above this temperature. (The asymptote shown for concentrations <10 ppm may be real, but more likely is a sampling and analysis problem.)

As can be seen from Figures 12-4 and 12-5, the rate of toluene destruction is very fast at temperatures greater than 1400°F. Conversions greater than 95% can be obtained with residence times of .1 second or less without any need for flame contact. However, several problems remain.

1. These data were taken in a plug flow reactor so all solvent was given equal treatment. There were no mixing delays and no cold spots.

2. The measurements were based on flame ionization detector readings (see Section 10.3) and only represent rates of hydrocarbon disappearance.

Figures 12-4 and 12-5 follow

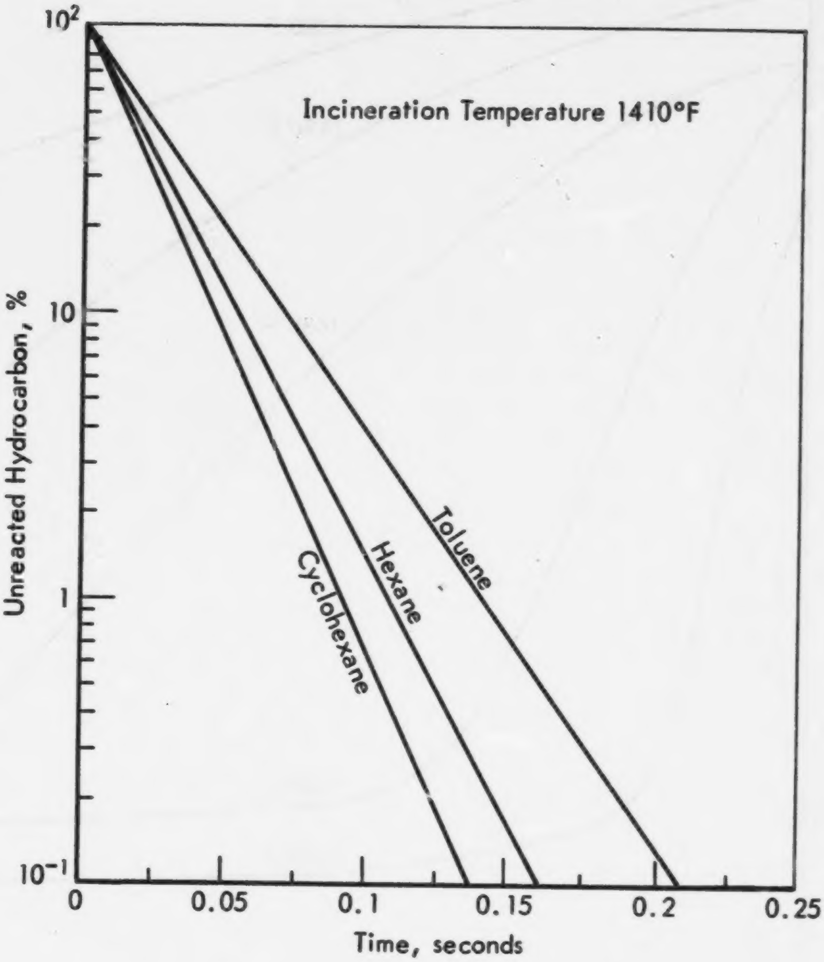

Figure 12-4. HYDROCARBON OXIDATION RATES IN ABSENCE OF FLAME

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& *The Heating and Ventilating Engineer Journal* 12-91

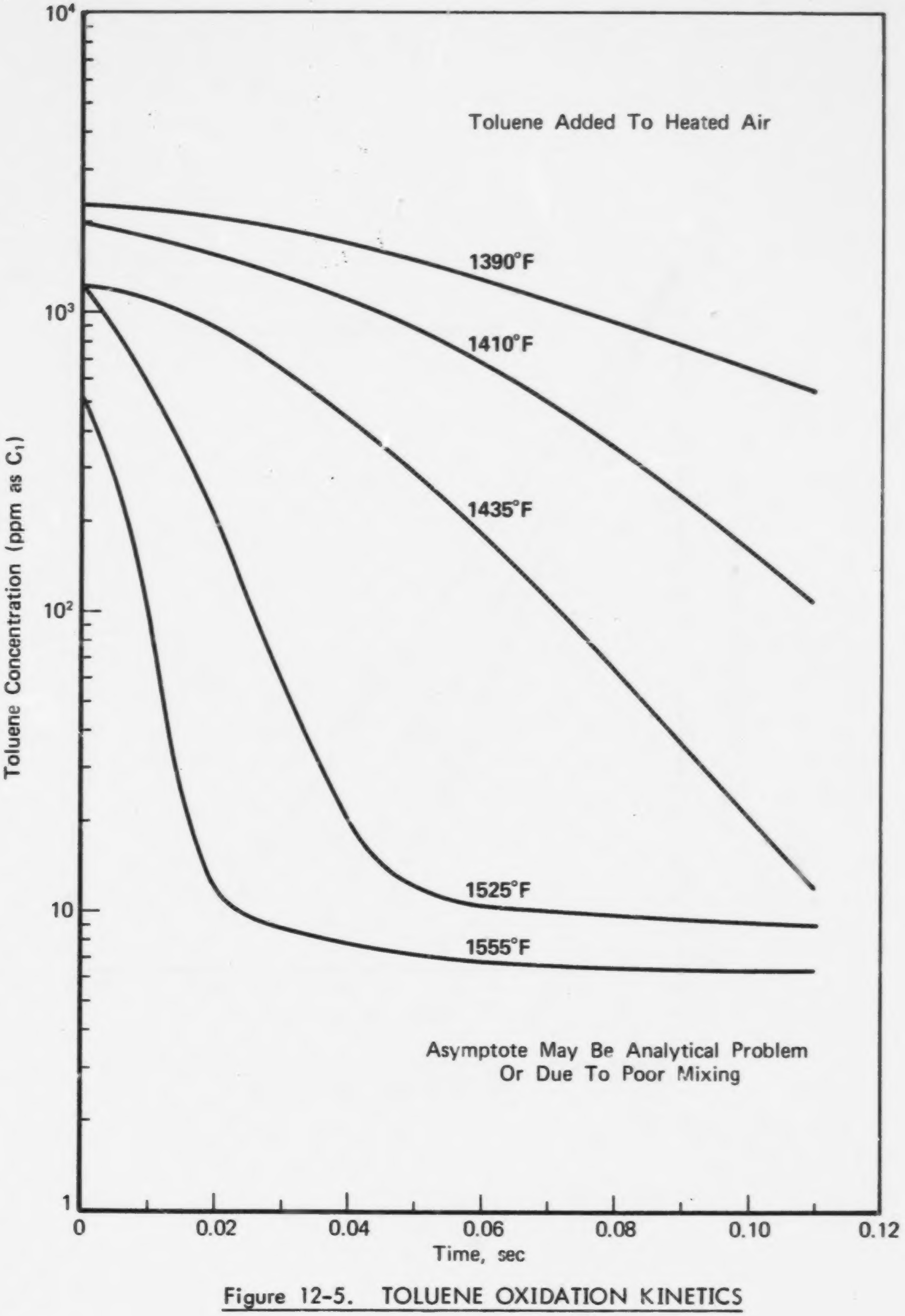

Figure 12-5. TOLUENE OXIDATION KINETICS

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& The Heating and Ventilating Engineer Journal (12-9)

Aldehydes and CO do not show up. At these temperatures aldehydes should be rapidly oxidized but CO conversion is rather slow. Figure 12-6 is an estimate of the equivalent first order oxidation of CO compared to toluene from Figure 12-4.

Methane is one of the most difficult hydrocarbons to oxidize either thermally or catalytically. Its rate of combustion has been measured many times but always at high concentrations and temperatures. An estimate for this rate is[12-10]

$$\frac{-d[CH_4]}{dt} = 4.7 \times 10^{18}\, e^{-56600/RT}\, X_{CH_4} X_{O_2}{}^{1/2} X_{H_2O}{}^{1/2} \left[\frac{P}{R'T}\right]^2 \text{ gmoles/cm sec} \qquad (12\text{-}29)$$

where P [=] atm
R' = 82.06 cm^{3}atm/gmole°K
R = 1.986 cal/gmole°K
T [=] °K
X_i = mole fraction of i

Using this equation, the extent of conversion versus temperature has been plotted in Figure 12-7 for 18% O_2 and 6% H_2O. It should be stressed that this equation has been extrapolated far below its intended range of methane concentration >5% and temperatures >2000°F. Again, there is only CH_4 oxidation to CO. Conversion of $CO \rightarrow CO_2$ is not included.

Methane oxidation rate can also be calculated from the rate of the reaction

$$CH_4 + OH \rightarrow CH_3 + H_2O \qquad (12\text{-}14)$$

This rate expression is

$$\frac{d[CH_4]}{dt} = k_{14}[OH][CH_4] \qquad (12\text{-}30)$$

As discussed in Section 12.2.3, radical concentrations in afterburners are close to equilibrium values. Therefore, the concentration of hydroxyl radicals [OH] can be taken at its equilibrium value for

$$O + O + M \rightleftarrows O_2 + M \qquad (12\text{-}13)$$

and

$$O + H_2O \rightleftarrows OH + OH \qquad (12\text{-}8)$$

This gives

$$n_{OH} = \frac{K_8{}^{1/2} n_{H_2O}{}^{1/2} n_{O_2}{}^{1/4}}{K_{13}{}^{1/4}(R'T)^{1/4}} \qquad (12\text{-}31)$$

where n_i = gmoles of component i
R' = 82.06 cm^3atm/gmole°K
T [=] °K
K_i = equilibrium constant for reaction 12-i

Taking the rate of reaction 12-14 as[12-11]

$$k_{14} = 50 \times 10^{-13} e^{-9900/RT} \ cm^3/gmole\ sec \qquad (12\text{-}32)$$

we get

$$\frac{d[CH_4]}{dt} = 2.82 \times 10^{15} e^{-47200/RT} [CH_4][O_2]^{1/4}[H_2O]^{1/2} \qquad (12\text{-}33)$$

where [ ] = concentration in gmoles/cm^3

Figure 12-7 also shows the extent of conversion versus temperature using this equation.

12.3.2 Carbon Monoxide Oxidation

As previously noted, the oxidation of $CO \rightarrow CO_2$ is usually the rate limiting step in premixed flames and is expected to be the slowest step in the destruction of gaseous pollutants in afterburners as well. There are a great variety of rate expressions given for this reaction, and these are discussed below. Afterburner experience shows that temperatures in excess of 1400°F are required for satisfactory CO cleanup in .5 seconds total residence time (including mixing time). Figure 12-6 compares the estimated rate of CO oxidation with that shown in Figure 12-4 for hydrocarbons. It is obvious that longer times and/or higher temperatures are needed when CO cleanup is desired rather than merely hydrocarbon destruction.

In the absence of water, CO is extremely difficult to burn. This is because no OH radicals can form to oxidize CO via reaction 12-4. Therefore, the much slower reaction of CO with O_2 is required. Afterburners are generally employed to clean up fume streams containing water vapor and solvents composed of hydrogen as well as carbon. In addition, the combustion of methane generates much water. Therefore, adequate water is always expected to be present.

As with methane, carbon monoxide oxidation kinetics have been determined by many investigators, but concentration and temperatures were much higher than those of interest in dilute fume incineration. The rate expressions developed

Figures 12-6 and 12-7 follow

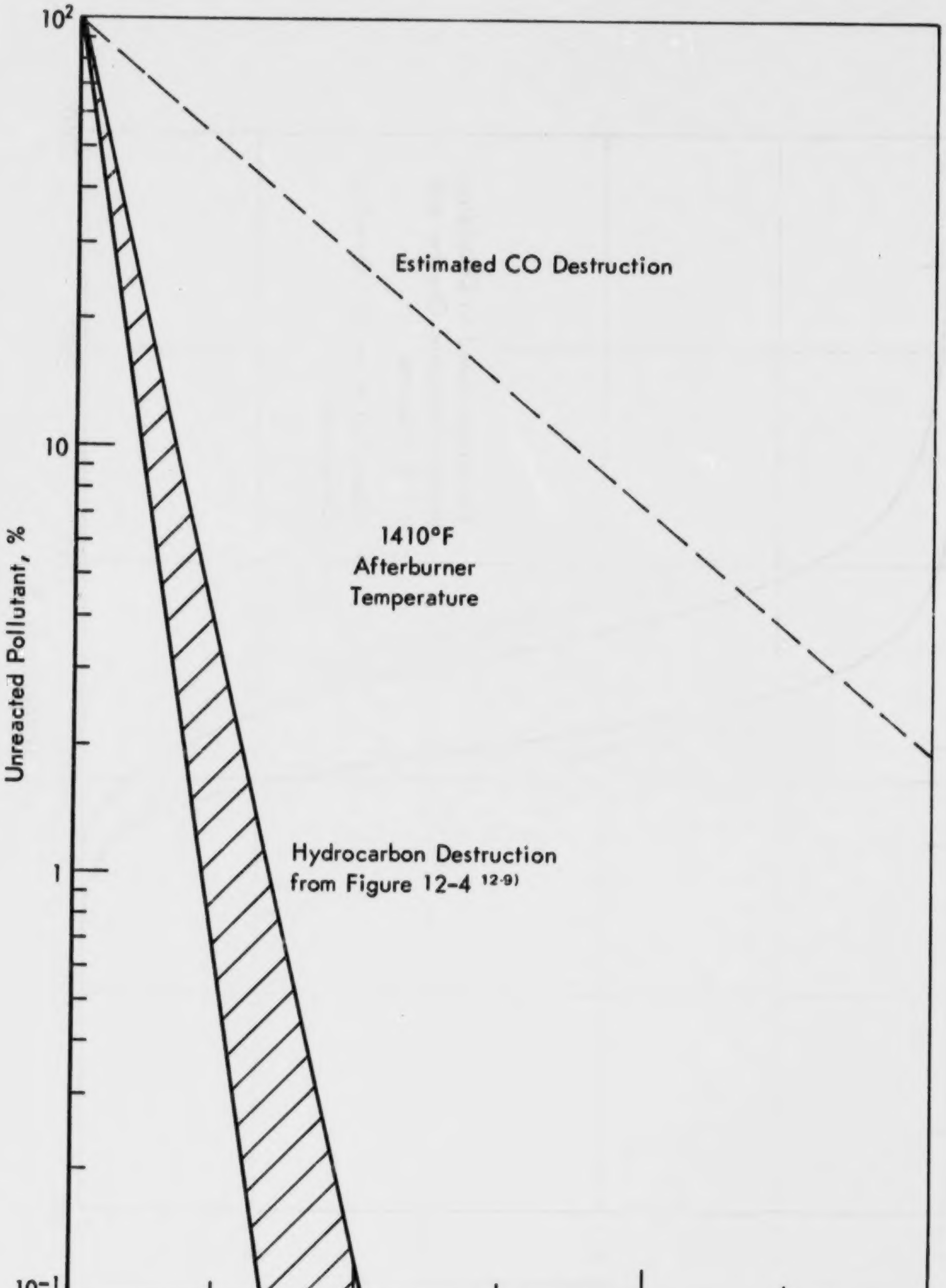

Figure 12-6. COMPARISON OF HYDROCARBON AND CARBON MONOXIDE DESTRUCTION RATES

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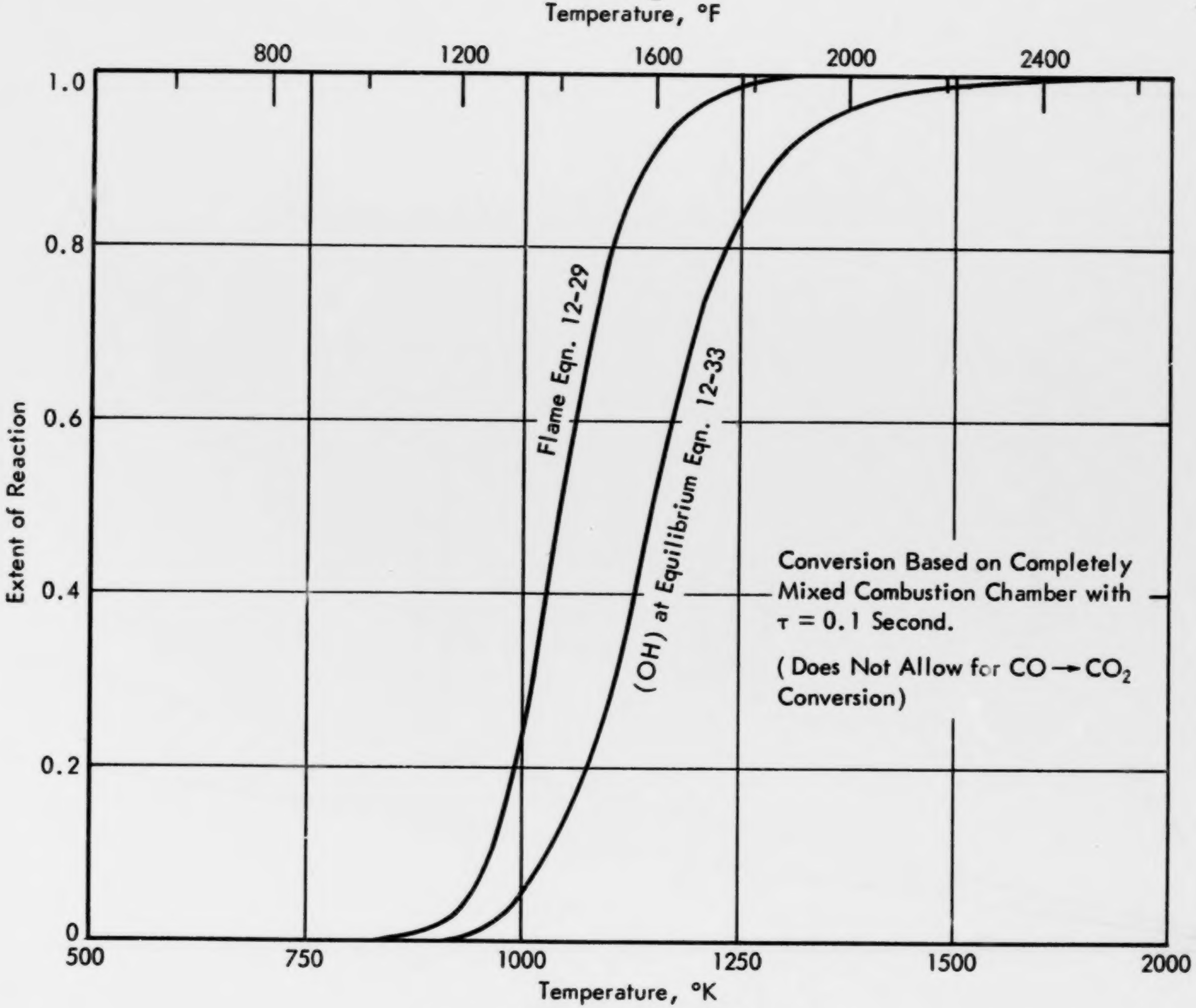

Figure 12-7. METHANE OXIDATION
Comparison of Rate in Flame with Rate Calculated for Equilibrium (OH)

in several of these flame studies are given below and these have been used to calculate the extent of CO oxidation at various temperatures for 6% water, 18% oxygen and .1 second residence time. These are plotted in Figure 12-8. Again, the rate expressions have been extrapolated far below their intended temperature and concentration range and should be suspect. As can be seen, there is some disagreement between the various expressions.

Field, et al.,[12-12)] surveyed several of the findings of investigations and give the expression

$$\frac{-d[CO]}{dt} = 3 \times 10^{10} X_{CO} X_{O_2}{}^{.3} X_{H_2O}{}^{.5} \left(\frac{P}{R'T}\right)^{1.8} e^{-16000/RT} \qquad (12\text{-}34)$$

where nomenclature is the same as for Equation 12-29

This low activation energy gives unrealistically high reaction rates at low temperatures.

Morgan[12-10)] also determined CO oxidation rates in his intensely mixed spherical reactor and gave the expression

$$\frac{-d[CO]}{dt} = 1.8 \times 10^{13} X_{CO} X_{O_2}{}^{.5} X_{H_2O}{}^{.5} \left(\frac{P}{R'T}\right)^{2} e^{-25000/RT} \text{ gmoles/cm}^3\text{sec} \qquad (12\text{-}35)$$

Koslov[12-8)] gives the expression

$$\frac{-d[CO]}{dt} = \frac{1.04 \times 10^{12} X_{CO} X_{O_2}{}^{.25} X_{H_2O}{}^{.5}}{T^{2.5}} e^{-32000/RT} \text{ gmoles/cm}^3\text{sec} \qquad (12\text{-}36)$$

Finally, Yuster, et al.,[12-5)] (the data we used in Section 12.1) give

$$\frac{-d[CO]}{dt} = 1.91 \times 10^{2} e^{-35600/RT} p_{O_2} p_{CO} \text{ lbmoles/in}^3\text{-sec} \qquad (12\text{-}37)$$

where p_i = partial pressure of component i, psia

Figure 12-8 also contains a plot of CO oxidation rate calculated assuming OH radicals are at their equilibrium concentration and the rate of Equation 12-5

$$CO + OH \rightarrow CO_2 + H \qquad (12\text{-}5)$$

is given by[12-11])

$$k_5 = .71 \times 10^{-13}\ e^{-7700/RT}\ \text{cm}^3/\text{gmole sec} \qquad (12\text{-}38)$$

Using the equilibrium concentration of OH radicals given by Equation 12-31, we get

$$\frac{-d[CO]}{dt} = 4 \times 10^{13}[CO][O_2]^{1/4}[H_2O]^{1/2} \qquad (12\text{-}39)$$

where concentrations are given as gmole/cm^3

It can be seen that this equation gives very much lower rates of CO oxidation than the flame determined data. Actually observed rates in afterburners tend to lie in between these two extremes. The CO oxidation rate shown in Figure 12-6 is representative of expected rates, and this is about halfway between the extent of reaction predicted by Equations 12-36 and 12-39.

Equation 12-39 has also been used to calculate the extent of reaction vs time for a constant temperature of 1250°K. The results for a plug flow and completely mixed combustion chamber are shown in Figure 12-9. This shows the importance of obtaining a high enough temperature since if backmixing is important, it is difficult to achieve high levels of conversion even for long residence times. Of course, the rate given by Equation 12-39 is unrealistically low, but a similar plot in practice would correspond to temperatures of 1300-1400°F.

All of the expressions reviewed above for the rates of oxidation of hydrocarbon and CO indicate a significant dependence of the oxidation rate on the oxygen content of the stream. In normal afterburner application, the waste stream is essentially contaminated air, with an oxygen content of 15-20%m even after passing through the afterburner. However, in some potential applications, the waste stream comprises combustible contaminants in an inert carrier (e.g. N_2 or CO_2); and air (as a source of oxygen) must be blended with the waste stream before passing through the afterburner. The amount of air added must be sufficient to bring the oxygen content of the stream being treated up to a reasonable level, if the oxidation rate is to be high enough for adequate conversion under normal afterburner temperature and residence time conditions. Some equipment manufacturers have specified 4 - 5%m O_2 as a minimum in the afterburner effluent. The kinetic expressions indicate that lower concentrations could be used, but at the expense of increasing temperature or residence time to achieve the necessary conversion.

Figures 12-8 and 12-9 follow

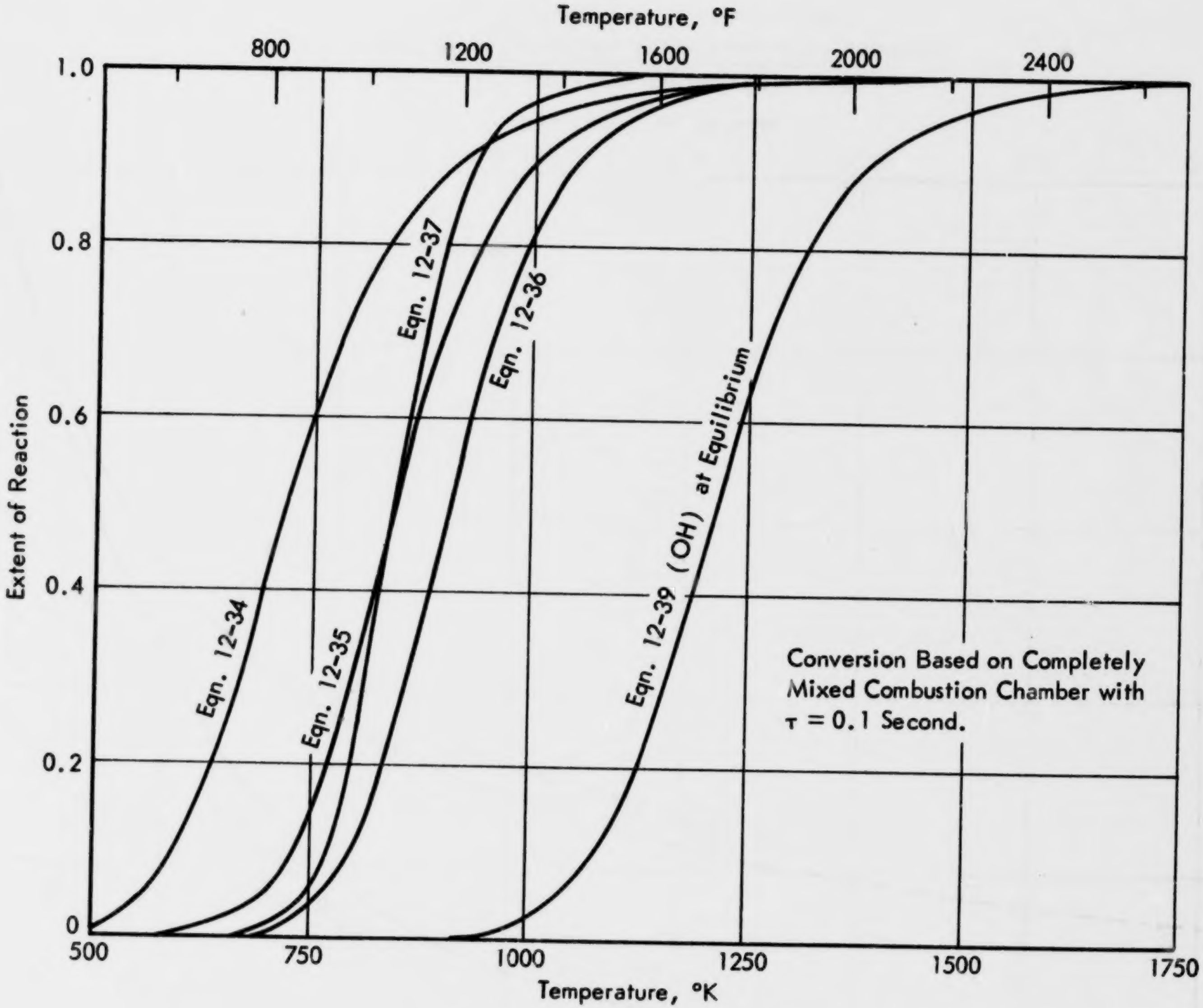

Figure 12-8. CO OXIDATION
Comparison of Flame Measured Rates with Rate
Calculated for Equilibrium OH Concentration

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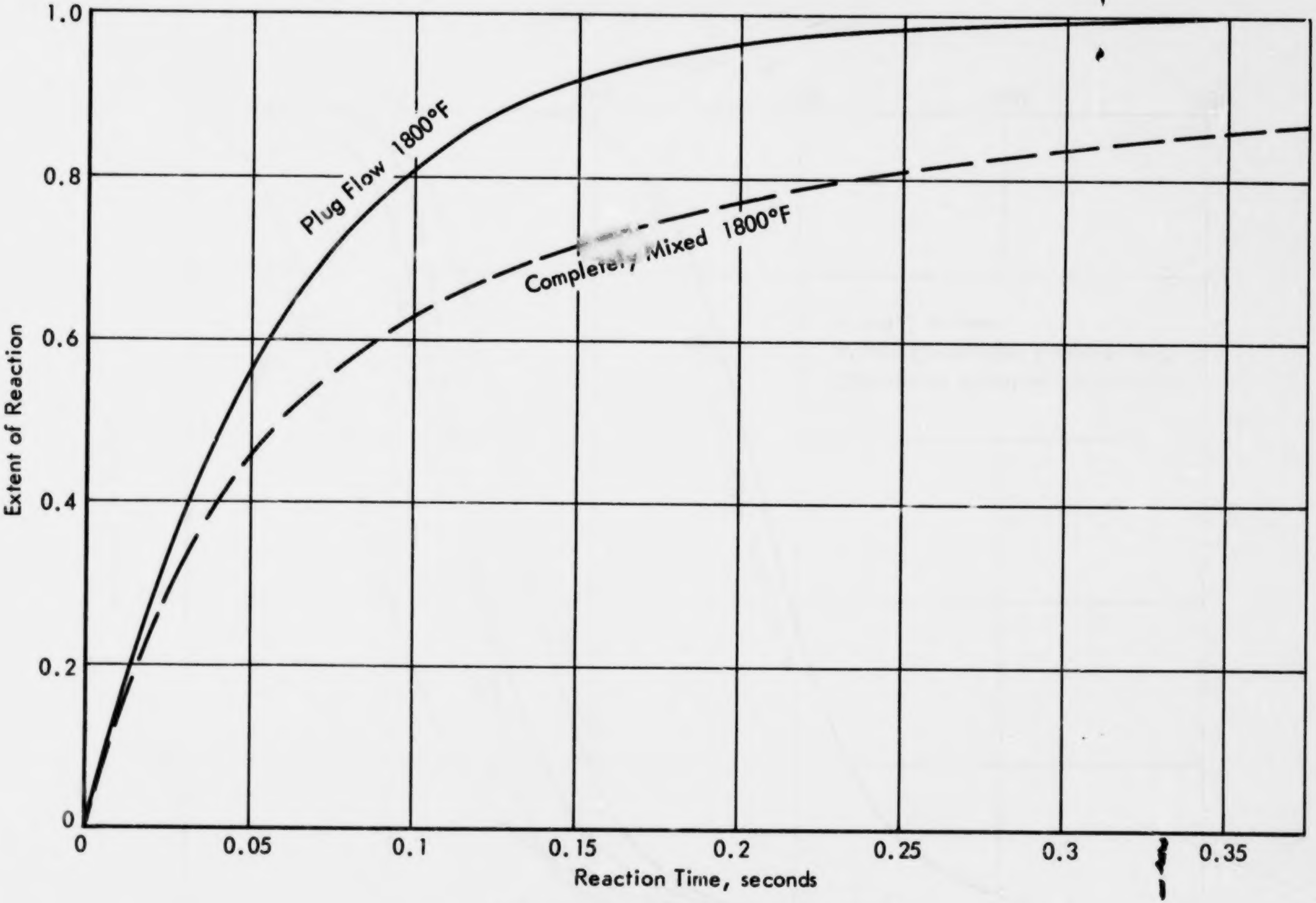

Figure 12-9. CO OXIDATION RATE
Plug Flow vs Completely Mixed Combustion Chamber.
Rate Calculated for Equilibrium (OH).

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12.3.3 Liquid Smokes and Droplets

Combustion of liquid smokes and droplets requires an additional step compared to gaseous pollutants since vaporization must precede oxidation. All organic "liquids" which are present in fume streams are expected to have boiling points below 600-700°F. Therefore, vaporization should take place rapidly at typical afterburner temperatures of ~1400°F. In fact, when smoke emissions are eliminated using afterburner temperatures of 800-900°F, vaporization is the major phenomenon occurring. The effluent is invisible but analysis would find high aldehyde and CO concentrations since oxidation is very slow at these temperatures.

Liquid fuel combustion involves sequential vaporization and burning in the gas phase so the problem has been extensively studied. An excellent review was recently published.12-13) The rate at which liquid droplets vaporize and burn is dependent on droplet size and the rate at which heat can be transferred to the surface. Rapid mixing and radiation of heat from the combustion chamber walls or flame give a high rate of heat transfer.

Time required for vaporization and destruction of original hydrocarbon can be estimated by12-14)

$$t = \frac{29800\ M_W d_o^2}{p_{O_2} T^{1.75}} \qquad (12\text{-}40)$$

where p_{O_2} = oxygen partial pressure, atm

d_o = original drop diameter, cm

T [=] °K

M_W = molecular weight

For a 200 molecular weight hydrocarbon, this formula predicts burning times at 1000°K (1340°F) of .0046 seconds for a 50 micron drop and .0185 seconds for a 100 micron drop. If temperature is raised to 2000°K (3140°F), which is a typical flame temperature, 100μ drops burn in .0055 seconds. These times are of the same order as are commonly reported for single droplet combustion. Afterburners should not have to handle droplets larger than 50 → 100μ since these are easily removed in simple cyclones. Therefore, the additional time required to burn up droplets rather than gases (<.02 seconds) is not an important consideration in sizing the afterburner.

12.3.4 Soot and Combustible Particulates

As contrasted to liquid smokes, solid carbon, soot and other combustible particulates are expected to increase the required afterburner temperature and residence time. For low temperatures (<2000°F) and small particulates (<100μ) the rate of burnup is controlled by the rate of reaction rather than diffusion. This is expressed as a surface reaction rate q [=] g/cm²sec. If one assumes spherical particles, this results in a linear burning rate variation with particle size

$$q = \left(\frac{1}{\pi d^2}\right) \frac{D(\rho_s \pi\ d^3)}{6\ Dt} = \frac{\rho_s}{2}\ \frac{Dd}{Dt} \qquad (12\text{-}41)$$

where D indicates differentiation

ρ_s = density of particle

d = particle diameter

Integration gives the time for complete burnout as

$$t = \frac{\rho_s d_o}{2q}$$

Field$^{12-12)}$ has reviewed data relating to the combustion of pulverized coal and coal char and found that most results can be correlated by

$$q = 8710\ e^{-35700/RT}\ p_{O_2} \qquad (12\text{-}42)$$

where q [=] g/cm^{2}sec
T [=] °K
p_{O_2} = partial pressure of oxygen, atm
R = 1.987 cal/gmole°K

This has been used by Niessen, et al.,[12-15] to predict burnup of smoke and char particles in incinerators. For oxygen partial pressures of .1 atmospheres, this equation predicts the burning times given below.

BURNING TIME FOR COAL CHAR

Temperature	Particle Size		
	100 μ(micron)	10μ	1μ
1460°F	4.3 seconds	.43 seconds	.043 seconds
2000°F	1.3 seconds	.13 seconds	.013 seconds
2540°F	.14 seconds	.014 seconds	.0014 seconds

The numbers in this table have been generated by extrapolating Equation 12-42 far below the temperatures and particle sizes used in experiments which were the basis for the equation. In general, particle sizes were in the range used in pulverized coal boilers (30-60μ) or larger (up to 2000μ). Temperatures were typical of flame combustion, 2000°F and higher. Therefore, the burning times should be taken as estimates only.

These char burnup rates do not appear to accurately reflect the rates of soot combustion. Lee, Thring, and Beer,[12-16] Fenimore and Jones,[12-17] and Magnussen[12-18] have measured the oxidation rates of soot formed in fuel-rich hydrocarbon flames. (Once formed, the soot is burned with an excess of air.) They find that the surface reaction rate, q, for soot is approximately two orders of magnitude below that given by Equation 12-42. The reasons for this may be the small size of soot particles, <400A (.04μ), which is of the order of the mean free path of the gas molecules; changes in surface properties (Lee, et al.,[12-16] reported only 50% more surface area than accounted for by the external surface of 400A spheres); or lack of any volatile matter. Pure graphite also gives very low burning rates.[12-19] Lee, et al.,[12-16] correlated their data according to

$$q = \frac{1.085 \times 10^4 p_{O_2}}{T^{1/2}} e^{-39300/RT} \qquad (12\text{-}43)$$

Again, temperatures were above 2000°F but extrapolation to our range of interest for afterburners give

SOOT BURNING RATE

(p_{O_2} = .1 atm)

Temperature	Particle Size		
	10μ	1μ	.1μ
1520°F	1750 seconds	175 seconds	17.5 seconds
1700°F	400 seconds	40 seconds	4.0 seconds
2060°F	43 seconds	4.3 seconds	.43 seconds
2600°F	4 seconds	.4 seconds	.04 seconds

Fenimore and Jones[12-17]) correlate their results in terms of collision frequency between OH radicals and soot particles. If OH concentration is taken as the equilibrium value for p_{O_2} = .1 atm and p_{H_2O} = .08 atm, one calculates soot burnup times roughly 1/2 of those given in the previous table. Magnussen's correlation[12-18]) agrees fairly well at temperatures above 2000°F, but predicts much lower rates than shown in the table at low temperatures. Oxygen concentrations of 21% will double the predicted rates.

This has serious consequences for afterburners designed to eliminate dense carbon smokes like soot. It may be necessary to use very high temperatures and long residence times to achieve burnup. Experience with wire burnoff, cloth carbonization, and smelting operations shows that temperatures of over 1700°F and residence times greater than .5 seconds are often needed to clean up the smoke plume. However, refuse incinerator smoke can often be eliminated at temperatures below 1500°F. The amount of volatiles present, particle size, and surface properties all help determine the burning rate.

12.3.5 Flame Inhibition and Combustion of Substituted Hydrocarbons

As previously mentioned (section 12.1), any diluent will reduce the rate of combustion by absorbing heat and lowering the concentration of the reacting species, thereby making it more difficult to maintain a flame. In this way, nitrogen, carbon dioxide, water (except when present in small amounts during combustion of hydrogen free fuels), and other inerts inhibit combustion. Since both reactant concentration and heat balance are affected, the effect on combustion rate is fairly strong, e.g., maximum flame speed for propane is decreased by a factor of 3 as the oxygen concentration in the enriched combustion air is halved from 40% to 20%.

Certain substances, however, exercise much stronger effects even when present at a very low concentration. Halogen and phosphorous compounds act in this way, and as a result, they are used as flame retardants and fire extinguishants. In fume incineration, these same substances sometimes must be burned.

The mechanism involved in this later form of flame inhibition is still not fully understood but it is thought to be primarily through interference in the free radical, chain branching mechanism. Recently, a literature review was published by Creitz[12-20)] giving complete coverage of the experimental findings and mechanistic interpretations of inhibition by halogen compounds. One of the more detailed analyses of the mechanism of flame inhibition was presented by Rosser, Wise, and Miller.[12-21)] They claim that most inhibitors first decompose to yield a halogen atom which forms the halogen acid by reacting with the fuel. The halogen acid then reacts (with) one of the chain branching species (H, OH, or O) to replace it with a halogen atom and prevent branching, e.g.,

$$CH_3Br \rightarrow CH_3 + Br \qquad (12\text{-}44)$$

$$CH_4 + Br \rightarrow CH_3 + HBr \qquad (12\text{-}45)$$

$$OH + HBr \rightarrow H_2O + Br \qquad (12\text{-}46)$$

This mechanism has not been proven but it does form the basis for most discussion. Fenimore and Jones[12-22)] consider the mechanism to be primarily through a reduction in concentration of H atoms.

$$H + CH_3Br \rightarrow CH_3 + HBr \qquad (12\text{-}47)$$

with no further effect assigned to the halogen acid. Inhibitors are more effective in rich flames than lean flames since the concentration of H atoms is high and hydrogen plays a more active role in the combustion process.[12-2)] In fuel lean flames, the concentration of H is low but reactions 12-6 → 12-9 are in local equilibrium so a reduction in H results in reduced OH and O as well.

Phosphorous compounds and various metal salts can also serve as flame inhibitors, probably again by interferring with chain branching mechanisms and serving as third bodies for recombination reactions. They are widely used as fire retardants in fabrics and paints. Also, high surface area in the combustion chamber will inhibit oxidation by speeding recombination reactions. If the inhibiting surface is also a heat sink, quenching is likely to occur.

Many studies have been conducted to determine the extent of inhibition for low concentrations of various halogens. This is reported as the flame speed for premixed flames at various air to fuel ratios with various amounts of inhibitor in the mixture. Flame speed or propagation velocity is a measure of the kinetics of the reaction but cannot be directly interpreted in terms of rates of reaction. It is measured[12-23)] by optically determining the surface area of the flame and dividing this into the volumetric flow rate of reactants. Typically, halogens show a 2-20% maximum reduction in methane flame speed per 0.1 mole percent of inhibitor added. Bromine is one of the most effective of the halogens. There is much scatter among the various investigators, at least in part because flame speed is dependent on the experimental set up, initial temperature, and gas flow rate.

Substituted hydrocarbons generally have a lower heat of combustion and are more difficult to burn. Methane has a heat of combustion of 23,875 Btu/lb, while that for CH_3Cl is 5850 Btu/lb and that for CH_3Br is only 3492. Methyl bromide is barely combustible with range of only ±1/2% of stoichiometric mixture. Certain wastes are not sufficiently combustible to support combustion alone and supplementary fuel must be added. $CHCl_3$ with a heat of combustion of 1345 Btu/lb will burn alone only in pure oxygen and even then, only under carefully controlled conditions.

A few studies have been conducted in which the flame speeds for pure halogen compounds were measured. In order to sustain a flame, oxygen concentrations greater than 21% must often be used. In pure oxygen, the flame speeds of the various chlorinated methane compounds compared to pure methane were ~1/3 for CH_3Cl, ~1/8 for CH_2Cl_2, and ~1/44 for $CHCl_3$. Combustion is inhibited both by interference in the mechanism and a lower rate of heat generation per mole of reaction.

From the above discussion one can get a feel for the fact that in general substituted hydrocarbons are more difficult to oxidize than unsubstituted or oxygenated species. An afterburner handling these species must allow for operation at a higher temperature and/or a longer residence time. Some preliminary test work should be conducted to determine design specifications.

12.3.6 Elementary Reactions - Kinetic Models

As noted in the previous sections kinetic data applicable to fume incineration are relatively limited. The overall rate expressions allow order of magnitude calculations but there is disagreement between various experiments. Also, temperatures and concentrations are far different than those of interest. Prediction and extrapolation can be done with more confidence by using a model based on the actual mechanisms involved in the combustion. This involves using the elementary reaction steps (free radical reactions) discussed in Section 12.2. Computer programs for integrating large systems of non-linear differential equations have been adapted to handle reaction rate expressions and many kinetic models of flames have been developed. These have been especially important in understanding nitrogen oxide formation in flames. Heat generation and simple mixing processes can be accommodated. This type of model cannot be used (given current knowledge) for a priori prediction of reaction rates but it holds much promise as a semi-empirical tool. Difficulties involved with a completely theoretical approach lie mainly in determining the reaction mechanism (to decide which elementary reactions should be included in the model), and in determining the reaction rate constants. Methods are available for estimating these rate constants for simple reactions (e.g., hydrogen abstraction) and many rates have been determined through mass spectrometer studies of flames or in non-flame experiments. Some data are given in References 12-25 and 12-26. (Reference 12-26 presents the results of using a kinetic model to calculate CO oxidation rates in an automobile exhaust reactor.) It is most likely, however, that this type of model will be utilized for interpreting experimental results and making them useful in design. In this way, a minimum amount of experimentation will be required and interpolation and extrapolation can be done with confidence.

12.4 Mixing Processes - Control of Overall Oxidation Rate

The oxidation rates discussed in Section 12.3 assume that the combustible component and oxidant are well mixed and in molecular contact at time zero. In most cases of fuel combustion, fuel and oxidant are not premixed and combustion rates are controlled by mixing rates. (At flame temperatures, the initial hydrocarbon will disappear in <1 millisecond and even CO oxidation only takes 5-10 milliseconds once mixing has occurred.) As discussed in Chapter 13 flame shapes can be predicted quite well for simple geometries by calculating the zone where mixing between fuel and air is just sufficient to produce a (stoichiometric) mixture. The chemical kinetics can be ignored. As is discussed in Chapter 14, mixing time can also be an overriding consideration in afterburner design. Poor mixing is often the reason for poor performance.

Mixing processes are discussed in detail in Chapters 13 and 14. However, it is of interest to estimate the effect of mixing time on the rate of combustion from observed versus theoretical combustion intensities as measured in Btu/ft^3-hr. It has been determined[12-27] based on bimolecular reaction theory that the maximum theoretical combustion rate for a stoichiometric mixture of butane and air with instantaneous mixing is ~730 x 10^6 Btu/ft^3-hr. This is for optimal addition of premixed reactants to the combustion zone which gives a temperature ~80% of the adiabatic flame temperature. The maximum combustion intensity which has been observed experimentally[12-1] is ~100 x 10^6 Btu/ft^3-hr using an intensely mixed (uniform temperature and concentration) spherical combustor. The highest rates attained in practice, ~10 x 10^6 Btu/ft^3-hr, are in highly efficient gas turbine combustors employing strong recirculation patterns (to transport heat and radicals to incoming reactants) and rapid mixing. Normal burners, even with some recirculation, usually exhibit combustion intensities of 0.5 x 10^6 Btu/ft^3-hr or less and most combustion chambers are designed for heat release rates an order of magnitude lower than this. Thus, we can see that mixing times appear to be more than two or three orders of magnitude slower than chemical kinetic times, at least at flame temperatures. Within an afterburner, times for mixing and complete oxidation are closer to the same order.

12.5 Flue Gas Composition

Given sufficient reaction time, equilibrium considerations will determine flue gas composition and may effect the choice of design variables and operating conditions such as temperature, amount of excess air, use of supplemental fuel. When dealing with pure hydrocarbons in the presence of slightly more than the stoichiometric amount of oxygen, equilibrium is essentially completely toward CO_2 and H_2O. Any presence of intermediates in the flue gas is due to quenching of the oxidation reactions and is due to kinetic rather than equilibrium considerations. Equilibrium considerations are much more important when considering the fate of halogens, phosphorous, metals, nitrogen, sulfur, or other non-hydrocarbon components introduced into the fume incinerator.

12.5.1 Determining Combustion Equilibrium

Equilibrium constants can be generated for most any desired reaction by using the thermochemical data contained in the JANAF Tables.(12-28) Among other data, one can obtain free energy data as a function of temperature for compounds with respect to their elements. This can be used to calculate equilibrium constants since

$$K_p = e^{-\Delta F/RT} \qquad (12\text{-}48)$$

where $\Delta F = \sum_i \nu_i F_i \Big|_{\text{products}} - \sum_i \nu_i F_i \Big|_{\text{reactants}}$ = change in free energy due to reaction

F_i = free energy of component i

ν_i = stoichiometric coefficient of component i

K_p = equilibrium constant at constant pressure

Much of the information in the JANAF Tables is also available on a magnetic tape for use with computer programs requiring thermochemical data. Computer programs are available which utilize this data to calculate equilibrium compositions for large numbers of components by minimizing total free energy for the system.

12.5.2 Fate of Heteroatoms

When the fume stream contains contaminants other than plain hydrocarbons or oxygenated species, the flue gas may require treatment besides incineration. This may entail wet scrubbers, cyclones, filters, or other particulates removal devices. In deciding whether or not such cleanup is required, one must determine the concentration and form in which heteroatoms and inert particulates will be found in the flue gas. Some guidance as to maximum acceptable concentrations was presented in Section 3.1.2.7. The presence of acids in the flue gas and combustion chamber will introduce special corrosion problems unless planned for in the design (see Sections 7.2 and 7.3).

When halogens are present in the waste to be burned, they can appear in the flue gas as either the corresponding acid or as free halogen. In most cases, it is desirable to maximize conversion to the acid, since it is more easily removed from the flue gas and results in fewer corrosion problems.

The equilibrium distribution between the halogen and its acid is determined by the following equation, shown for Cl_2/HCl.

$$2\,HCl + 1/2\,O_2 \rightleftarrows Cl_2 + H_2O \qquad (12\text{-}49)$$

Hydrochloric acid is favored by low levels of excess air, high levels of water vapor (a high H/Cl ratio in the total feed to the incinerator, including supplemental fuel and steam is equivalent), and high temperature since the equilibrium constant shifts toward HCl as temperature increases. (K_p for reaction 12-49 is ~2.0 at 1000°F, ~.1 at 1700°F, and ~.02 at 2400°F.) At 2000°F and 25% excess air with a 2:1 H/Cl mole ratio in the feed, the equilibrium ratio of Cl_2/HCl is .016. If the temperature is dropped to 1700°F this ratio increases to ~.04. If the H/Cl ratio is increased to 2.5:1, the Cl_2/HCl ratio drops to .011 at 2000°F. Incineration of dilute fume containing some halogenated solvents involves high levels of excess air and low temperatures and probably results in the formation of considerable free halogen. Equilibrium ratios of Br_2/HBr are two orders of magnitude greater than Cl_2/HCl under similar conditions. Efficient combustion is essential when handling halogen substituted organics to avoid formation of poisonous components such as phosgene, $COCl_2$. This is especially a danger when burning at low temperatures with insufficient oxygen in the burning region.

Organophosphous compounds break down to give P_2O_5 in the flue gas. This has a very low vapor pressure below ~700°F and forms submicron hygroscopic solid particles below this temperature. These exit as a phosphoric acid mist when water is present and result in a highly visible plume, unless concentrations are below 1 mg/SCF. Corrosion of metallic components and refractory can be severe unless proper materials of construction have been utilized.

Metals which are present in the waste stream are most likely to be converted to the corresponding oxide if the incineration temperature is high enough. (Inert particulate matter is most likely already present as an oxide and will not be changed.) Some of the metal oxides will react with water to form hydroxides and will be found in the flue gas as a fine mist. Sodium and other alkali metals usually appear as hydroxides. Whatever form the metals take in the flue gas, they can be expected to be highly visible, submicron drops or particles.

Sulfur and nitrogen are other components likely to be found in wastes. The fate of sulfur in flames has been extensively studied since it is present in fuel oil and is the cause of our sulfur oxides pollution problems. In an oxidizing flame, the sulfur is primarily converted to SO_2 with 10% or less appearing as SO_3. Nitrogen oxide formation is a problem whenever combustion occurs since some atmospheric nitrogen is oxidized along with much of the nitrogen in the fuel. Afterburners do not seem to be a significant source of nitrogen oxides due to the relatively low temperatures at which they operate. Below 2000°F very little NO_x (indicating various oxides of nitrogen) is formed due to low oxidation rates. Only in the flame is there much of a problem and rapid mixing with cold fume limits the amount which can be formed. Most of the chemically-bound nitrogen in the waste can be expected to form NO along with that formed by nitrogen fixation in the burner flame. When burning natural gas in the preheat burner, less than 100 ppm NO_x would be expected to be formed due to nitrogen fixation in the primary flame products from the burner. Subsequent dilution of these flame products with the fume stream being treated would reduce this NO_x concentration to 25-30 ppm or less in the afterburner effluent.

Chapter 13. BURNERS

13.1 Introduction

Chapter 3 has given general description, performance, and selection information on flame-type afterburner systems. Chapter 12 immediately preceeding has described the general principles of combustion, mixing, and retention at temperature which apply to the flame-type afterburner. The following section describes the physical equipment in which the operating principles are carried out. It is intended to give more detail and a broader basis for understanding that lies behind the description of Chapter 3.

In this chapter the convenient breakdown into separate steps, as shown in Figure 12-3, will be adhered to. That is, the burner, the subsequent mixing of burner combustion products and fume stream, and the retention at an elevated temperature for a specified time, will be discussed as separate phenomena. In varying degrees, depending upon the actual afterburner design, these functions are intermixed and it becomes very difficult to say where combustion is complete, where the mixing function begins and ends, or where the beginning and ending of the reaction that consumes the fume pollutants takes place.

13.2 Equipment Types and Alternatives

Although this chapter is titled "Design Fundamentals", no attempt will be made to instruct on how to design a burner. There are a great variety of commercial burners on the market. The objective here will be to give some understanding of how the different burners work, and what their capabilities and limitations are. When selected and installed so as to match their capabilities and accommodate their limitations, these burners will give good service.

13.2.1 Function

The function of a burner is primarily to provide a source of heat. This function is initiated by bringing together a fuel source and an oxygen source at proportionate rates which will support combustion reaction in the form of a flame. Usually this ratio is proportioned to give a small excess of oxygen over that required for a chemically complete conversion of the fuel to combustion products. The chemically correct ratio is called the "stoichiometric" ratio. The mixture containing more air than the stoichiometric ratio requires is sometimes called "fuel-lean"; that containing less than the required air is "fuel-rich".

The burner must be designed so that the flame is retained in a fixed position with respect to the burner, with a predetermined flame pattern. It should have an ability to retain the flame in position over a range of firing rates.

A secondary function of the burner is that its flame serves as a source of free radicals, which are highly reactive but transitory molecular species that can attack the organic components of the fume stream and initiate their being broken down and oxidized. The "half life" of these free radicals is generally very short compared to the time required to intimately mix combustion products from the burner with fume introduced external to the burner; nevertheless, they may play an important role in decomposing the fume.

13.2.2 Discrete Versus Distributed-Source

In a discrete source burner, as shown in Figure 13-2, a single large flame plume is formed into which the fume stream must ultimately be blended. A distributed source burner, on the other hand, Figure 13-3, gives an array of small flames. Fume and flame are already subdivided, thus substantially reducing the space required for mixing in the fume stream. If one can assign a "characteristic diameter" to the flame plume, as shown in the sketch, and describe a length, L, which is required in order to get a completion of the combustion process and a mixing-in of the fume stream, then the ratio L/D will be a characteristic number for describing the process, whatever the scale. This L/D applies of course only if the flame patterns are similar, and ignores the presence of nearby walls, which can effect the shape of the flame appreciably. One can further assume that the flow velocities are about equivalent, since they are governed externally by the pressure drop available from the blower and the fuel supply. Clearly the distributed source burner completes the combustion reaction and subsequent mixing-in of fume streams in a considerably shorter distance and time scale than the single source burner can do. The distributed source design is particularly applicable where gas fuels are available and where the fume can serve as the oxygen supply for the burner.

13.2.3 Premix Versus Diffusion Mix Burner

13.2.3.1 Premix

When gaseous fuels are available at elevated pressure, usually above 2 psi, premix burners may be used. Combustion air and fuel are mixed in a jet nozzle-mixing device before passing into the combustion space. The porting which admits the mixture to the combustion chamber are designed to give a fairly high velocity through a number of small orifices to avoid the possibility of having the flame "flash back" through the entering pipe to the mixing chamber. Figure 13-4 is an example of a premix burner. The arrangement shown would also give a distributed flame, but fume admitted around the burner must have a low velocity to avoid blowing the flame out. Porting may be distributed over the length of a pipe channel, giving a "line" or a "ribbon" burner, or it may be laid out in grid or wheel-and-spoke patterns. The kitchen gas range is a common example of the premix burner. The premix burner is usually found on very small afterburners with heat releases below 200,000 Btu/hour, although the size is not limited. Unless completely free from the possibility of fouling, the premix burner should not use the fume stream as an oxygen supply, since the deposit of any material would be sure to plug the small orifice openings that serve the burner as a flame arrester. This burner may also be operated with only part of the air introduced in the premix section, requiring that additional oxygen be supplied by the fume stream above the burner. It will give about a ten-to-one turndown ratio; the jet induced air supply remains in nearly constant ratio to the fuel

Figures 13-1 through 13-5 follow

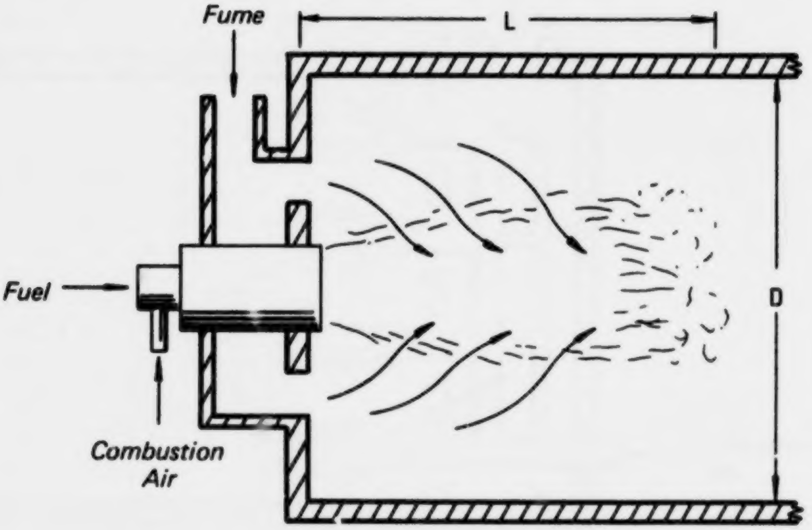

Fume
L
Fuel
D
Combustion Air

Figure 13-2. DISCRETE-SOURCE BURNER

206b

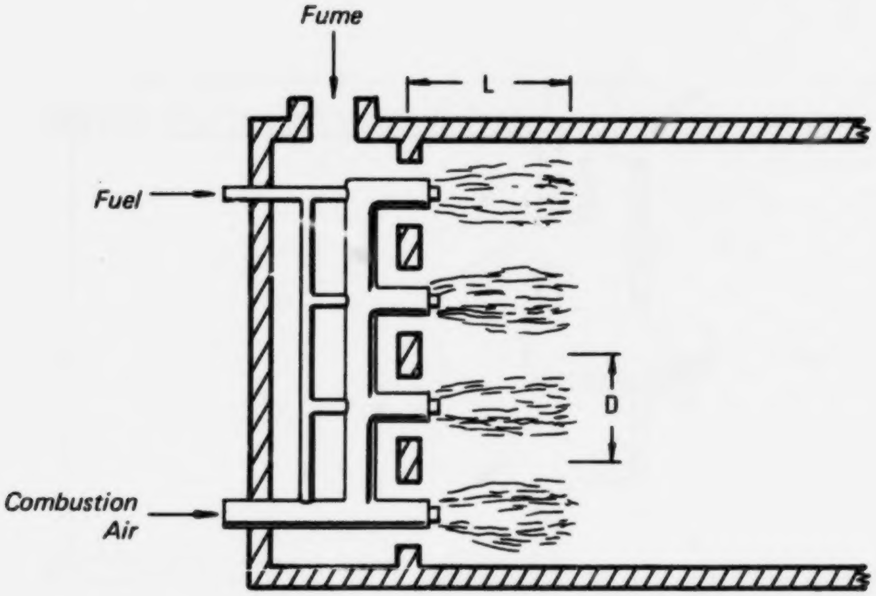

Fume
L
Fuel
D
Combustion Air

Figure 13-3. DISTRIBUTED-SOURCE BURNER

2066

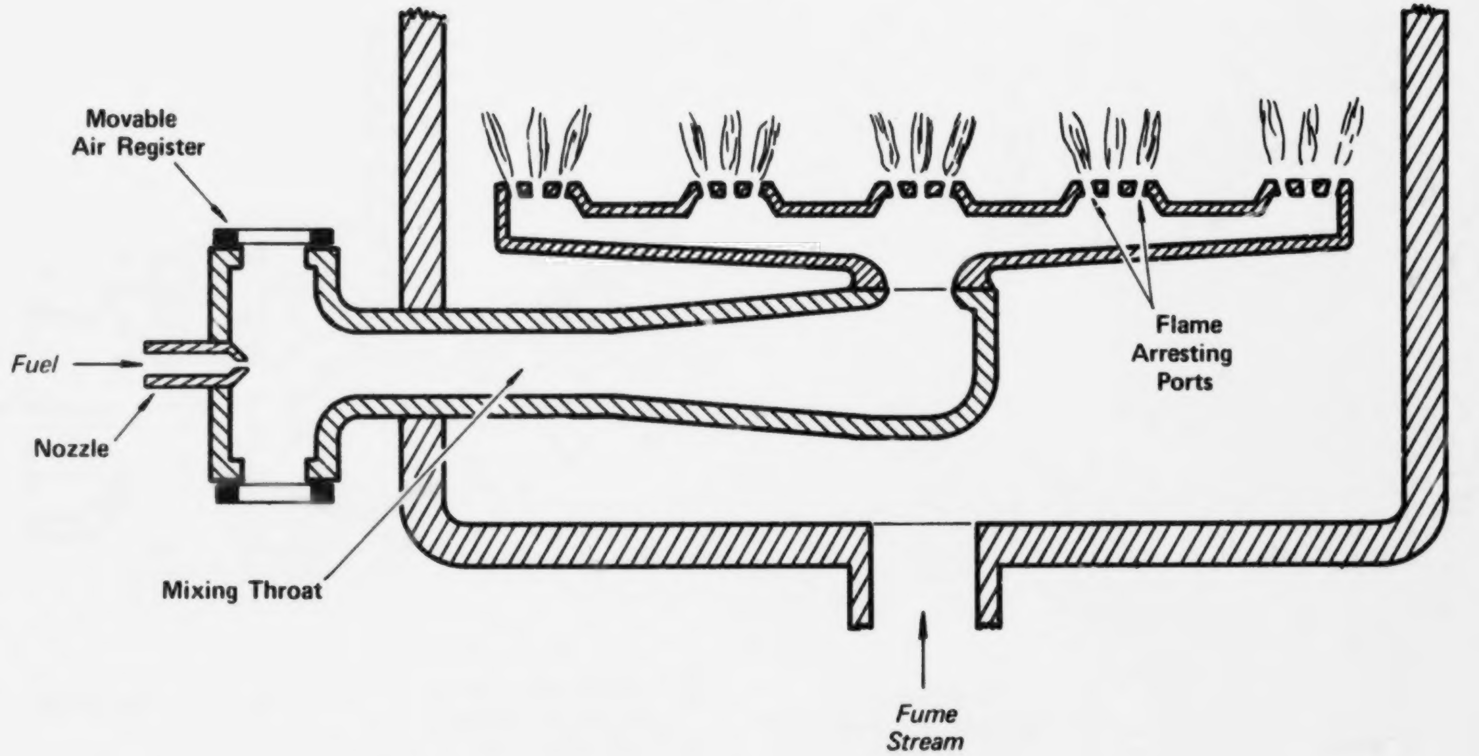

Movable
Air Register
Fuel
Nozzle
Mixing Throat
Flame
Arresting
Ports
Fume
Stream

Figure 13-4. PREMIX BURNER

S-14121
67784

S-14121
67784

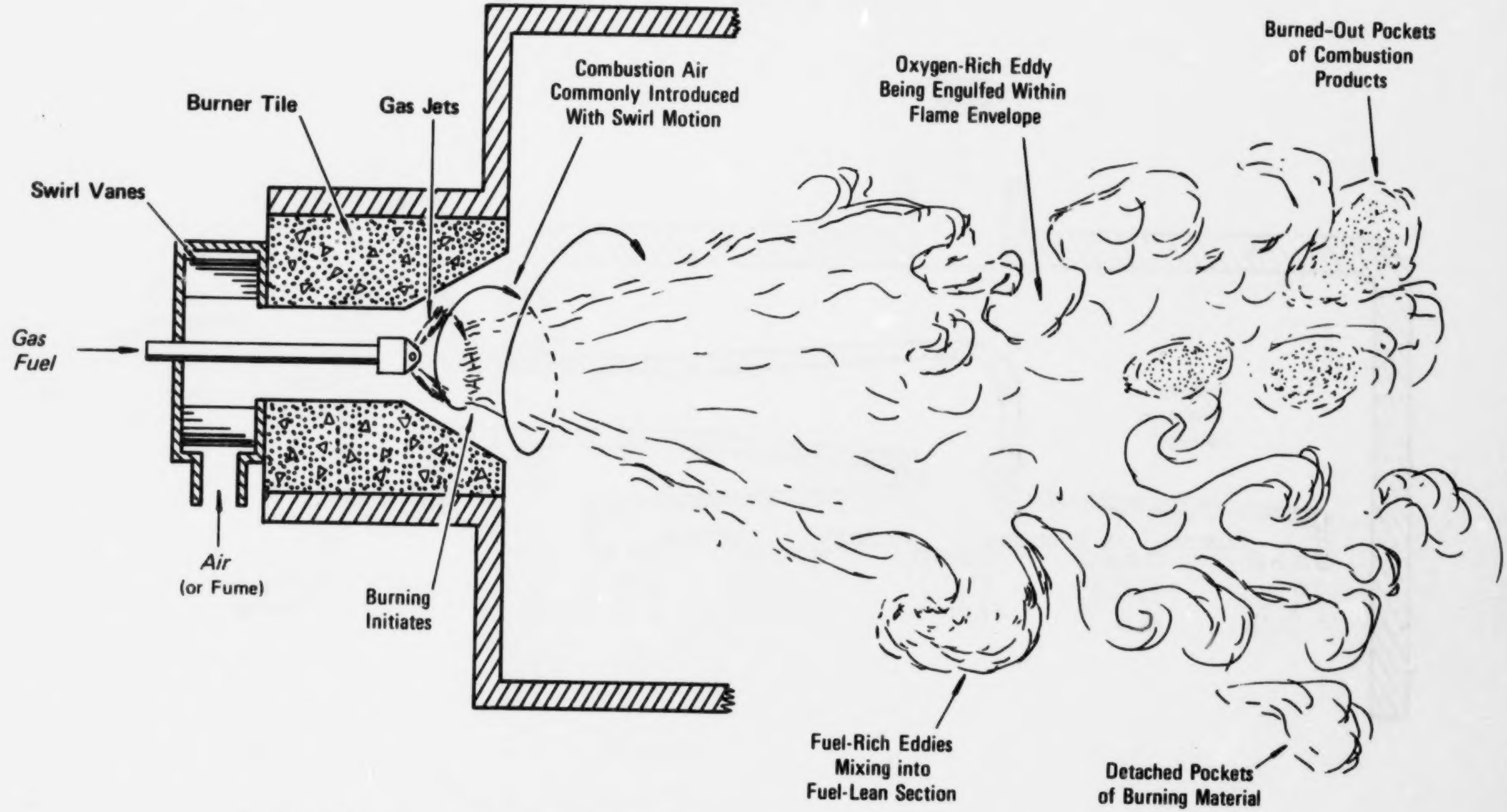

Swirl Vanes
Burner Tile
Gas Jets
Combustion Air Commonly Introduced With Swirl Motion
Oxygen-Rich Eddy Being Engulfed Within Flame Envelope
Burned-Out Pockets of Combustion Products
Gas Fuel
Air (or Fume)
Burning Initiates
Fuel-Rich Eddies Mixing into Fuel-Lean Section
Detached Pockets of Burning Material

Figure 13-5. SCHEMATIC REPRESENTATION OF A DIFFUSION-FLAME GAS BURNER

supply. Using external combustion air in this way, the burner is not bothered by having foul material in the fume stream, but this is done at the expense of poorer fuel economy.

13.2.3.2 Diffusion-Mixed

Alternative to the premix burner is the diffusion-mixed type, also called the nozzle-mixed type. The majority of burners found in afterburner service are of this type. Raw gas or atomized oil droplets are mixed with the air supply by a turbulent diffusion process occurring within the plume of flame. Figure 13-5 is a schematic representation of a diffusion flame gas burner with this mixing process taking place. Air (or fume) is introduced through a register that can control the amount passing through, depending upon the pressure available; often the register is in the form of vanes which will give the combustion air a swirling motion. Gas jets are introduced by a nozzle or a distributor ring placed in the air entry channel, with the jets disposed for mixing into the air. However, it is only some distance downstream from the gas jets where the velocity has slowed and where enough mixture within the combustible limits has been formed to support a flame. It is only within restricted limits of an air to fuel ratio that the burning will persist. The actual burning process is an extremely complicated one, and there are many chemical and physical processes occurring simultaneously. In the diffusion flame, the rate controlling process is that of the mixing between air and fuel, which involves a transport of both mass and momentum between regions of high velocity and regions of low velocity. Only in a relatively thin interface is there a combustible mixture, and this governs the formation of a sheet of flame. The sheet is very much warped and wrinkled as eddies of fuel are swept out into the air section, or as eddies of oxygen-containing air move into the fuel. Pockets of flame will be carried into unignited regions, which will burst into new flames; or they will be projected into zones that have already been burned out, whereupon the flame is diluted and quenched. Since the flame front cannot move at very high velocity, it remains always detached from the gas nozzle. A flame front will sweep through a stagnant mixture of cold gases at relatively low velocity, of the order of 2 to 30 feet/second. At too high a velocity the flame will actually blow out and away, and the flame will become extinguished; special provisions must therefore be made to hold the flame in place, as discussed below under "flame stability".

Some nozzle-mix burners distribute the gas supply through a series of jets or from a pipe ring with many openings into the throat section of the air inlet. In this way the air and fuel mixing are well under way before entering the combustion space. Such a burner is not clearly a premix or a discrete diffusion-mixed type.

The size and shape of the flame is largely determined by the means used for mixing gas and fuel, and for preheating it to ignition temperature. The resulting flame patterns differ widely from one burner to the next. This will be a topic of a later section.

13.2.4 Oil Versus Gas Fuels

The burning of oil further complicates the combustion process. Liquids do not burn (unless they are self-reactive materials), and it is only after a droplet has been formed, an envelope of evaporated combustible vapor has developed around the droplets, and some of this vapor has mixed with the air to form a combustible mixture, that a flame can take place.

Figure 13-6 gives a schematic representation of liquid flame phenomena. Depending upon the type of atomizer used the spray pattern can take a number of forms. Shown is a mechanically atomized spray, formed by feeding the oil under pressure through a nozzle. The liquid leaves the nozzle in the form of a thin liquid sheet, conical in shape, which then proceeds to break up into filaments, which further subdivide into tiny droplets. As a droplet passes out into the flame zone, it is heated by both radiation from the flame and by convection from the hot gases that surround it, and an evelope of vapor forms around it. The vapor in turn mixes with the surrounding air and becomes ignited. Depending upon the fuel used, a droplet may completely evaporate or it may reach a point where it chemically cracks to form burnable vapors and residual carbon char. The charred remains of droplets take the form of tiny hollow spheres, called "cenospheres". If the combustion chamber is hot enough and there is oxygen available, the cenospheres will be consumed; otherwise, they may leave the combustion chamber and be carried out the stack as unburned material. When burning heavy residual oils, droplets tend to be large (because of the high viscosity) and the flame pattern approximates a cloud of tiny projectiles trailing flames. With more volatile liquids, and with a fine degree of atomization, droplets are easily vaporized, and the flame becomes essentially the same as that of the diffusion mixed gas burner.

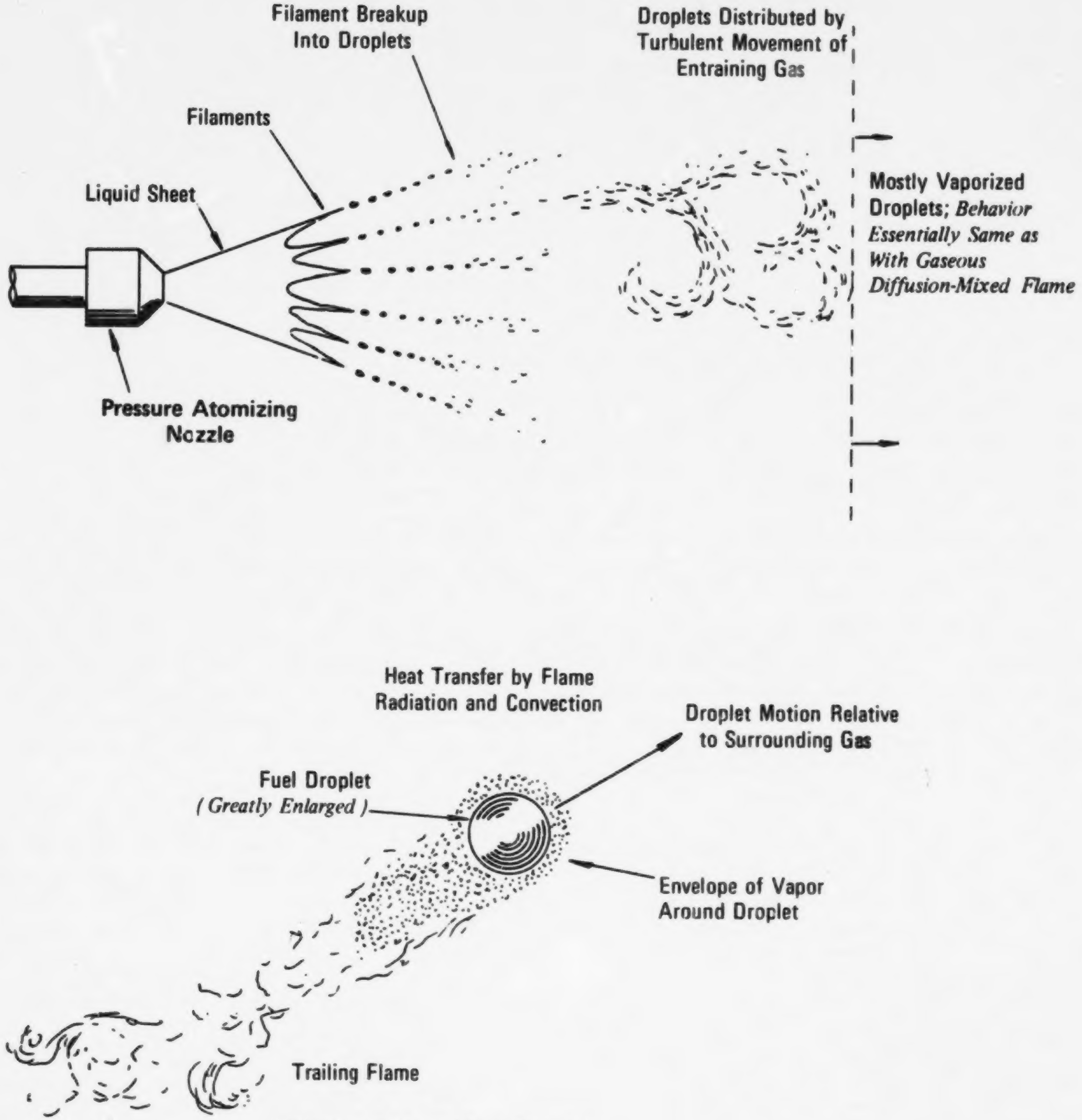
As with the gas flame, there are many patterns of flame possible with the oil burner. These depend partly upon the spray pattern of the atomizing system, and also upon the manner of introducing the combustion air, its axial and tangential or swirl momentum. Because of the problems of oil atomization (discussed in the next section) oil burners for afterburner service are discrete burners, used singly or at most there may be two or three burner units.

Oil and gas may be handled in the same burner, with provisions for withdrawing the oil burner gun with the atomizing nozzle when it is not in use (otherwise heating the nozzle tip when no fuel is flowing through it causes the fuel contained inside it to crack to carbon char and plug the small openings).

Gas fuel is expensive in some parts of the country and available only on an interruptable basis in other parts. For this reason many afterburners are designed with provisions for both gas and oil burning.

Oil burners often generate substantially higher levels of nitrogen oxides than do gas burner units, although the amount depends on the size and operating conditions of the burner, and the presence of bound nitrogen in the oil. In areas where nitrogen oxides are an important source of air pollution, the use of oil burners may be restricted.

Figure 13-6 follows


Filament Breakup
Into Droplets
Droplets Distributed by
Turbulent Movement of
Entraining Gas
Filaments
Liquid Sheet
Mostly Vaporized
Droplets; *Behavior*
Essentially Same as
With Gaseous
Diffusion-Mixed Flame
Pressure Atomizing
Nozzle
Heat Transfer by Flame
Radiation and Convection
Droplet Motion Relative
to Surrounding Gas
Fuel Droplet
(Greatly Enlarged)
Envelope of Vapor
Around Droplet
Trailing Flame

Behavior of Individual Droplet

Figure 13-6. LIQUID FUEL FLAMES

13.2.5 Oil Atomization

Pressure atomization is the most commonly used method, particularly for the lighter oils and the smaller sizes. In the usual pressure atomizing nozzle, the oil enters a small cup through tangential orifices, which cause the oil to swirl at high velocity in the cup. The outlet forms a dam around the open end of the cup, and the oil spills out over the dam in the form of a thin, conical sheet, which subsequently breaks up into filaments and droplets. The jet orifices are small, and dirt or solids in the fuel will plug them. The pressure drop through the jet orifices varies as the square of the oil velocity, and at reduced flow rates the atomization is poor. The turndown with pressure atomization is commonly about 4 to 1. Operating pressures run from about 25 to several hundred lb/in^2. Oil viscosity must be maintained within prescribed limits, usually under 600 SSU, and oil heaters may be necessary to do this.

Steam and compressed air are alternative drive media to give the necessary energy for fuel atomization. Steam under pressure is a preferred medium. Under the right conditions, steam reacts with soot and carbon monoxide formed in the flame, and helps to give clean combustion without smoke. Depending upon the viscosity of the oil fuel and on the design of the atomizing nozzle, the amount of steam used may run from as little as 5% to as much as 30% of the mass rate of fuel. Since about 1100 Btu/pound are required to vaporize the steam, the latter figure represents a large loss of heat energy. Compressed air is also preferred method for atomization if a dependable source of compressed air is available; however, one can rarely justify air compression just for atomization. Other burners use a rotating cup; oil is slung as a sheet from the lip of the cup across the air channel. This arrangement is not commonly used for afterburners because of difficulty in retracting the rotary cup mechanism when it is desired to convert the unit to gas fired operation. Other burner units use high pressure steam or air to generate trains of shock waves through which the oil jet is passed; the shock waves break the oil jet into an atomized spray. It is claimed that these sonic atomizers successfully atomize heavy pitch materials.

In general, oil burners are less desirable than gas burners because distributed burner arrangements are not practicable and because the oil burner tips require frequent maintenance and cleaning. Improved serviceability of oil burner gun equipment can be obtained by using automatic retractors, which will retract the guns from the hot part of the burner as soon as the oil flow is cut off; these devices are expensive.

13.2.6 Outside Air Versus Fume Oxygen Supply

There is a considerable economic incentive for using the fume as a supply of oxygen for the combustion process, if at all possible. Outside air brought in for combustion must be heated along with the fume stream to the afterburner exit temperature of 1400 or 1500°F. Figures 11-10 and 11-11 permit comparison of the gas fuel required with fume air and with outside air for combustion. The saving in fuel by using the oxygen in the fume stream without additional air is about 30% for a 1400°F afterburner. (This saving in fuel is translated into costs in Chapter 11.) The principal limitation to doing this comes from the possibility of fouling the burner passageways with deposited material. Where mildly fouling conditions are anticipated, it may be possible to use the fume in a discrete burner since the burner passageways are relatively large. Where fouling conditions are very heavy, it becomes necessary to use outside air rather than the fume stream, since the fume stream can then be brought into the afterburner through a single register opening of relatively large size, and made accessible for cleaning without disturbing the burner.

13.2.7 Flame Stability and Turndown Ratios

Holding a flame under a variety of operating conditions and firing rates depends upon maintaining a localized zone within or adjacent to the normal flame envelope which will always lie within the combustion limits for the fuel and oxygen mixture, and in holding the zone where the fluid velocities are not sufficiently high to sweep the flame away. The following are examples of ways that are used to bring about improved stability.

13.2.7.1 Burner Tile (Figure 13-7)

A conically shaped ceramic tile piece surrounds the burner opening. The diverging flow area slows the air velocity. The tile is heated by exposure to the flame to a glowing red temperature, and it re-radiates and conducts to the incoming mixture. In addition, its conical or cup shape, combined with a tangential swirl to the incoming air or fume stream may cause a recirculation zone in the flame center directly opposite the fuel nozzle which brings hot combustion products and radicals back into contact with the fresh fuel and air mixture for ignition.

13.2.7.2 Target Rod and Plate (Figures 13-7b and f)

The flame is made to impinge on a cylindrical rod or a flat plate (made of ceramic or temperature-resistant alloy). Eddies form on the downstream side of this target, and these contain pockets of fuel-rich and air-rich mixtures. Portions of these are swept off into the stream, forming multiple ignition sources; portions are retained and regenerated on the back of the target. The retained flame serves as a continuous pilot.

13.2.7.3 Fuel-Rich, Low Velocity Pocket (Figure 13-7c)

The pocket is located downstream from the main fuel addition, and has its own fuel supply. The low velocity in the pocket allows a flame to be retained, and to act as a continuous pilot for the main stream. John Zink Company uses a ledge in the wall of the throat of one of its gas burners in this manner.

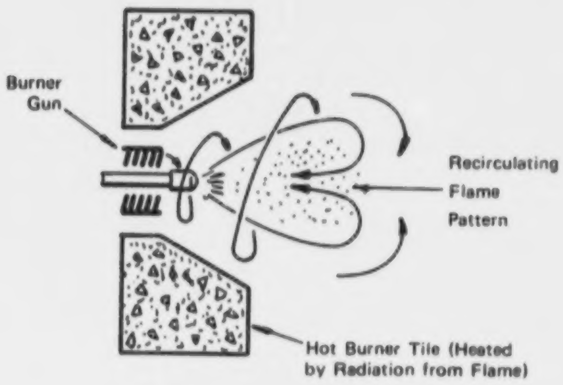
13.2.7.4 Graded Air Entry (Figures 13-7d and e)

The gas jet enters by a channel having a diverging cross section; air is introduced by holes that produce cross-stream jets of air. The diverging channel slows the mixture; meanwhile, some of the air jets are too fuel-rich, some too fuel lean; at some point in the channel they are near optimum. Wake eddies are formed that retain the flame. The Maxon Combustifume burner is of this type.

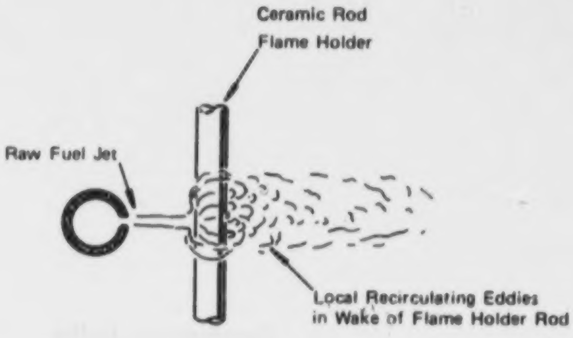
13.2.7.5 Deflector Plate on Fuel Gun (Figure 13-8)

A vaned conical plate, attached to the end of the fuel gun serves as a baffle and swirl-inducing device, holding a low velocity turbulent wake zone immediately opposite the burner tip. Although the mixture in this region is

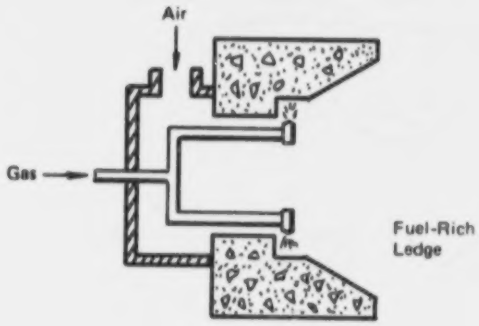
Figures 13-7 and 13-8 follow


Burner Gun
Recirculating Flame Pattern
Hot Burner Tile (Heated by Radiation from Flame)

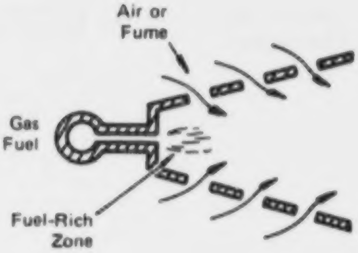
A. Recirculating Flame and Hot Refractory Tile


Ceramic Rod Flame Holder
Raw Fuel Jet
Local Recirculating Eddies in Wake of Flame Holder Rod

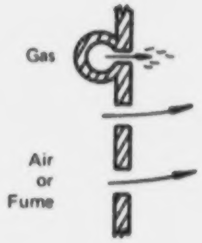
B. Target Rod


Air
Gas
Fuel-Rich Ledge

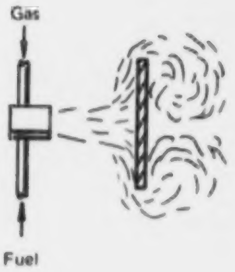
C. Flame Holder Ledge (After John Zink)


Air or Fume
Gas Fuel
Fuel-Rich Zone

D. Graded Air Entry Maxon Combustifume


Gas
Air or Fume

E. Graded Air Entry (Grid Plate Version)



F. Target Plate

Figure 13-7. EXAMPLES OF FLAME HOLDING ARRANGEMENTS

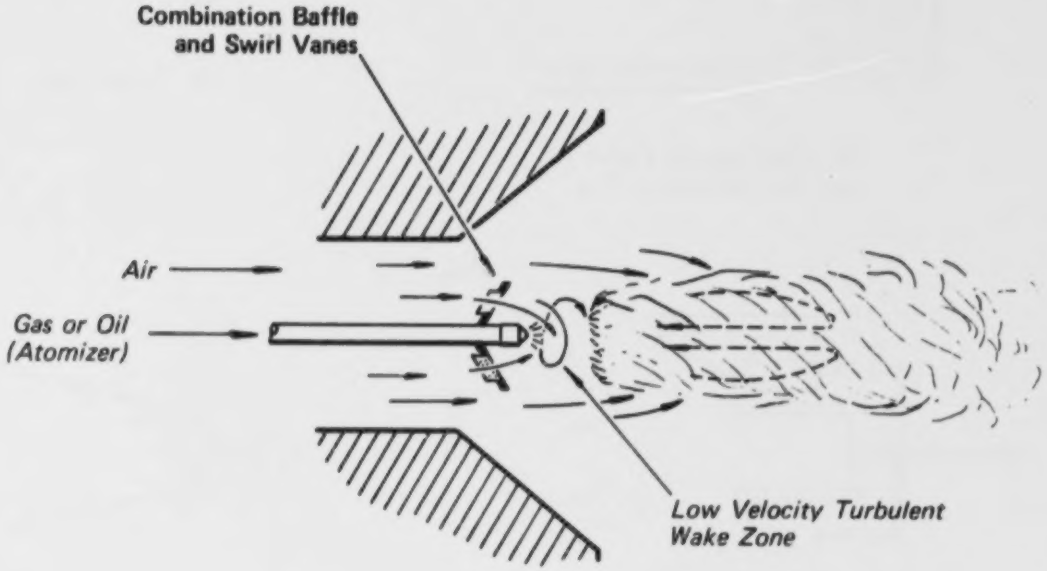

Combination Baffle and Swirl Vanes
Air
Gas or Oil (Atomizer)
Low Velocity Turbulent Wake Zone

Figure 13-8. FLOW PATTERN, AXIAL JET BURNER CONICAL OIL FUEL SPRAY NOZZLE

210C

generally fuel-rich, pockets of burnable mixture recirculate in the zone to hold the flame in place. Additional air sweeps past the plate undisturbed, and mixes with the central wake zone further out from the burner. This arrangement is especially common with packaged pressure-atomized oil burner units.

13.2.7.6 Preheated Air

A preheated air or fume stream used for combustion will improve the stability of the flame and increases the turndown range; the higher the temperature, the greater the effect. Preheat of the fume stream will result from the discharge from an oven, or from the use of heat recovery. When using heat exchangers a reduced flowrate through the exchanger will result in an increase in the temperature of the fume to the afterburner, provided the afterburner exhaust temperature is held constant; this may partly compensate for a limited turndown ratio rating of a burner that is based on ambient air ratings.

13.2.7.7 Holding Oil Flames

Flame holders of the types described are mostly applicable to gas burners. Retention of a flame in an oil burner is more difficult, since the complications of atomizing and vaporizing the fuel prevent introducing target plates, or generating small pockets of slow-moving mixtures. Flame retention is partly accomplished through using the hot burner tile and the recirculating eddy opposite the nozzle (Figures 13-7a) and aided by adding an air deflector plate near the nozzle tip (Figure 13-8). Sometimes the air supply is introduced in two sections, annularly disposed about the fuel nozzle. By inserting turning vanes, the air can be made to enter with varying degrees of swirl, which can promote mixing and burner stability (mostly this approach is used to control the shape of the flame as discussed below).

13.2.7.8 Excess Air Burners

Stability of the flame also depends on the achievement of near-stoichiometric mixture ratios of fuel and air (or fume), and most burners are designed to take from 5 to 40 percent excess air through the burner. The afterburner, as mentioned earlier, operates as an excess air device, in that sufficient fuel is added to heat the fume stream to about 1400°F (which is well below the stoichiometric flame temperature of about 3300°F). Thus, the afterburner (without preheat or heat recovery) runs about 250% excess air. Most burners will not tolerate putting the entire air or fume stream through the air register, since it will extinguish the flame. However, some burners have been specially designed as "excess air burners" for air-heating, oven-heating, and afterburner use, and the additional air can be tolerated. Generally, these burners are of the diffusion-mixed type, arranged so that the excess air is introduced into the outer part of the flame envelope, or downstream of the flame-proper. There is a danger that such a burner will quench the flame before the burning is complete, forming aldehydes, carbon monoxide, and other pollutants, which must then subsequently be destroyed in the afterburner holding section. Provided satisfactory performance in this respect is demonstrated, this type of burner avoids the separate step of blending in the bypassed fume, which is accomplished only with difficulty. (See the following section.)

Another stability phenomenon that may be encountered is the generation of some form of acoustic resonance, coupled with a cyclic rate of energy release by the flame. A number of different conditions can give rise to this "feedback" effect. Often the effect, which manifests itself as a loud musical note, can be controlled by adjustments to the burner. Other cases prove much more difficult to control and require changes to the burner or its control system.

Burners that operate at high velocities have an additional stability problem. An instantaneous interruption in the fuel supply (as with a slug of water in the oil supply) is more likely to extinguish the flame than in a slow, low-intensity flame, since the residence time in the flame is shorter, and the recirculating eddies are removed more quickly and are not available to re-ignite the flame when the fuel supply is resumed. The flame-out detector will then shut the burner down, since a continued flow of fuel into a hot combustion chamber might ignite explosively and destroy the afterburner.

13.2.8 Turndown Ratios

The forgoing discussion shows that special provisions are needed to obtain a burner which will operate stably over a wide range of flow rates and with a broad range of air-to-fuel ratios. The simple diffusion-flame gas or oil burner usually has a turndown ratio (maximum to minimum fuel rate) of about 4 to 1. With the flame-holder provisions described above, turndown ratios of 20 to 1 or more are possible. Where this is important, the requirements should be specified at the time of purchase of the burner. Some afterburner units receive a nearly constant fume flow, and the only variation in burner heat release is that required to maintain operating temperature in the holding section. Other units may be called upon to accept a variable fume feed rate or fume composition (that is, with combustible components), and an adequate turndown ratio in the burner is imperative. Sometimes this can be accomplished by using multiple burners, which may be cut in or out as the load demands.

13.2.9 Susceptibility to Fouling

Fouling conditions can arise from a number of sources:

13.2.9.1 Vapors Carried in the Fume Stream

Vapor components will tend to condense on cold surfaces. Cold ductwork walls, fuel distributor manifolds, and burner passages in which the fume stream is accelerated, are likely points. For water-based condensibles, these can also be points of corrosion. Water and some oily materials can be drained away and disposed of separately. Other oily materials, most resinous materials, and most greases, will tend to form polymerizing layers that can plug passageways in the burner.

13.2.9.2 Entrained Droplets

These tend to collect on narrow passageways, baffles, distributor plates, turning vanes, grid plates, etc.--wherever a flowing stream changes direction, or a stagnant gas pocket can form. As with condensed vapor materials, the liquid may be drained away harmlessly, or there may be a buildup of layers of solids on the surfaces. Since the mechanism of deposit is by impingement on the surface, deposits may accumulate in small openings in the burner, and char to form a hard coke. They may also constitute a fire hazard, and initiate a fire in the ductwork leading to the afterburner.

13.2.9.3 Entrained Particulates

Solid particles will drop out of the stream when the velocity is low, or impinge on surfaces where the flowing stream changes direction. They will collect in the wake of brackets, flanges, baffles, etc. If combustible they can be a fire and possibly an explosion hazard. If sticky, they are likely to coat all passageways, and block burner passageways.

13.2.9.4 Combinations

Most fouling materials are likely to contain all the fouling constituents: Condensible vapors, resinous (and polimerizable) droplets, and solid particulates. The afterburner purchaser should have a clear idea of what kind of fouling conditions he may experience when he approaches an afterburner manufacturer, and how frequently he is willing to shut the unit down for cleanout.

For gas-fired burners, where fouling is not anticipated, distributed burners using the oxygen available in the fume stream are recommended. For oil burners in clean service, likewise, the fume stream should be used for combustion air.

Applications of this class would be:

Most odor destruction units
Paint bake ovens
Resin curing ovens (depends on resin)

When light fouling is anticipated, use of distributed burners and the fume oxygen may be possible, but should be reviewed carefully. Burner passages for the fume stream should avoid openings less than 3/4-inch, and should be accessible for cleaning. Operating schedules should permit time for regular cleaning. It may be necessary to lag cold gas manifolds to avoid condensing vapors, and to insulate walls and ducts exposed to cold ambient conditions.

Applications of this class would be:

Resin curing ovens (depends on resin)

Under conditions of heavy fouling discrete burner units using outside combustion air should be used.

Applications of this class would be:

Rendering plants
Varnish cookers
Asphalt blowing units

Some manufacturers have made special burners intended to operate under severe fouling conditions. Fume passages are made without abrupt changes in direction, and with smooth channel walls. Fuel nozzles project into the stream with a minimum of frontal area, and with a streamlined shape. Such a burner would be used only under special conditions, and would be designed and built specially for the application.

13.3 Flame Patterns

The size and shape of the flame, or flame pattern, will vary greatly with the type of burner and the way it is operated. Not so immediately apparent, but fully as important to the afterburner, are the velocities, the axial and rotational momentum of the gas stream, the intensity of the eddy mixing process within and at the boundary of the flame, the concentration profiles and temperature profiles in the cross-section of the flame and around it. Moreover, the flame interacts strongly with the combustion chamber that surrounds it; the gross flow patterns affect the mixing-in of furnace gases and fume with the flame and the formation of recirculating eddies; the hot refractory radiates to the flame and affects its temperature and energy release rate. Thus, the pattern of the burner is not unique to the burner, but depends also on the conditions of its operation and installation.

Predicting the resulting shape of the flame is largely the result of art and empiricism. A start has been made on representing the process mathematically, but only the simplest of geometries have been handled, and with rather arbitrary assumptions on many of the intermediate processes involved. Figure 13-9 gives an example of a mathematical solution of the temperature and fuel concentration profiles in a single-jet diffusion burner in a cylindrical furnace, as calculated on a digital computer. This approach may prove very useful in the future, but is not an easily accessible design tool.

What follows, then, is a pragmatic and approximate guide to the flame patterns that can be expected, which must be established as a basis for mixing of flame and fume. This mixing should be completed before specifying the retention time under "plug flow reactor" conditions, during which destruction of the fume is assumed to take place. Published data on actual flame dimensions are almost non-existent; what there are apply only to specific burners. In the absence of either a general analytical approach or of extensive data, we can only suggest that the afterburner designer apply to the burner manufacturer for recommendations on flame dimensions for his burner for the specific application.

13.3.1 The Premix Burner

The flame pattern for the premix gas burner can be treated as a simple jet of length which is ten times the diameter of the port opening. This approximation is crude, since the turbulence, the composition of the fuel, the air-to-fuel ratio, the degree of preheating, all effect this jet; mostly they shorten it. Since the premix burner is used in multiple port arrangements with small port openings, the flame will in any case be completed within a short distance of the burner. It is blue in color, non-radiant.

13.3.2 The Discrete Diffusion Burner

The range of flame patterns obtainable with this class of burner is very large. For a given rate of heat release, the flame can be a compact, high-intensity brush, or a large, billowing, "lazy" flame; it can be a pale bluish color, nearly non-luminous, a brilliant radiant yellow-white, or a smoky orange; it can flare out in a shallow, saucer-like cone, form a bushy ellipsoid, or be concentrated in a long, narrow pencil-like flame. The principal variables controlling this variety of flames are the type of fuel used, the amount of turbulence generated in the mixing between fuel and air, and the amount of axial and rotational momentum different parts of the fuel and air are given as they leave the burner. The most certain guide for any given burner is to obtain the prediction and recommendation of the manufacturer. Not all of the possible flame patterns will be useful for the afterburner, but some idea of the choices is desirable. The controlling variables will be discussed here in a qualitative way, with some approximate guides on the size of flame.

13.3.2.1 Effect of the Type of Fuel Used

Natural gas flames tend to be of low luminosity, especially when there is good mixing between the gas and the fuel, and the air supplied is only slightly in excess of that required for complete combustion. An increase in the air supplied makes the flame more compact, turn a hard, bluish white, and give a noticeable roaring noise. A decrease in the air supplied makes the flame longer, paler, and "lazier". A luminous flame can be gained by separating the air supply into two parts, with about a third of the air well mixed into the gas, and the remainder introduced at low velocity in the outer annulus. The incomplete combustion of the first stage forms soot, which gives the luminosity.

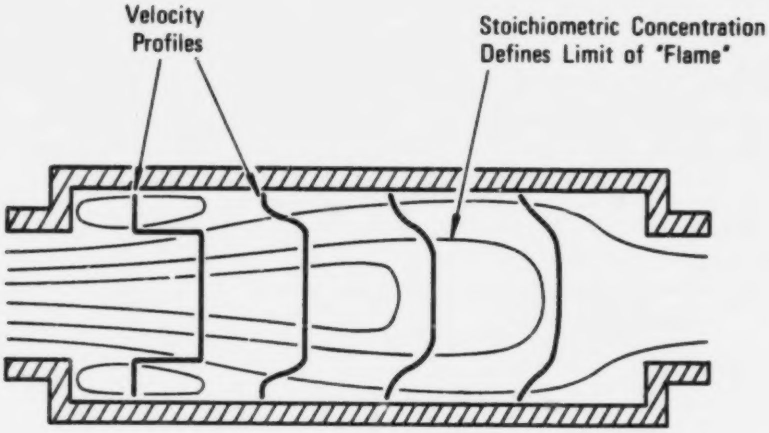
Unsaturated hydrocarbon materials present in the gas increase the amount of soot formed in the flame and make the flame luminous. Ethylene, propylene, butenes, butadienes, etc., are typical components involved.

Oil flames are generally luminous (exception: see paragraph below, describing the effect of using a pre-combustion chamber). Luminosity increases with the "heavy" oils, which usually have less hydrogen, and tend to form more soot particles, which become heated to incandescence. Because of the added time required to evaporate the fuel droplets and to mix the vapor with air, and because of the relatively slower mass transfer between the air and the soot particles, the flame is longer than with gas. (Also the gas flame is usually partly premixed.) If poor atomization results in forming large droplets, the flame may be very much longer, noticeably streaky, with the formation of particles of char that may not be fully burned.

When the flame of either a gas or an oil fuel is discharged into a hot refractory-lined chamber, it burns more quickly (i.e., making a shorter flame), and with somewhat less luminosity than when discharged into a cooled metal chamber (which tends to cool the flame and form more soot). When fuel, fume, or combustion air is supplied hot, the effect is to shorten the flame and usually to decrease its luminosity.

Figure 13-9 follows

Fuel-Air Reaction is Assumed Instantaneous at Point Where Stoichiometric Ratio is Reached as a Result of Turbulent-Diffusion Mixing.


Velocity Profiles
Stoichiometric Concentration Defines Limit of "Flame"

From Printout of D.B. Spaulding Program

Figure 13-9. CURVES SHOWING CONTOURS-FUEL CONCENTRATION AND VELOCITY

13.3.2.2 The Effect of Turbulence

Turbulence gives rise to the eddies which are the means for mixing the fuel gas or vapor in with the air or fume stream. Some turbulence is generated within the flame itself, but this is probably secondary to the effects of shear forces generated by the gross velocity differences across the flame. The source of this is the kinetic energy available from the gas and air streams. It is directly reflected from the available pressure drop supplied the burner. Since the mass rate of the air is some 15 times that of the fuel, unless a high gas pressure is available, most of this energy comes from the air stream.

In a gas burner much depends on the manner in which the gas is distributed into the air (or fume) stream in the burner. Some burners provide much better mixing inside the burner, leaving less to be done in the combustion chamber. This can be done by arranging multiple ports for releasing the gas into the air, with parts directing jets of gas across the air stream. Units with integral blowers will sometimes introduce the gas ahead of the blower, so that the blower blades aid in premixing the gas into the air. (The mixture must then pass into the combustion chamber with a velocity higher than that of flame travel, so that the flame will not flash back through the burner.) One design, the "Fanmix", introduces gas through channels in the rotating fan blades, obtaining very good distribution with relatively low pressure loss to the air stream. The better this initial distribution, the more closely the flame pattern becomes that of the "premix" burner; however, on this scale, the 10:1 L/D ratio between flame length and port diameter mentioned earlier would be excessive, and a perhaps 3:1 is more realistic.

In an oil burner this premixing is not possible, and the turbulence of the air stream and the passage of the fuel spray must be relied upon, taking their effects in the open combustion chamber.

The turbulence so far discussed, and its effects, are important, but subservient to the gross movements of the flame interacting with the rest of the gases in the combustion chamber; these gross movements result in important patterns of recirculation, and depend on the axial and rotational momentum of the gases as they enter the chamber.

13.3.2.3 The Effects of Axial and Rotational Momentum

The mass flow rate of the fuel and the air (or fume) entering the chamber with a finite velocity give the flame a substantial momentum. (Momentum = mass x velocity; it is a vector quantity, since it must be specified both as to magnitude and direction). If the stream enters by way of a straight tube, the momentum is axial, and the flame has the form of a straight jet. If the stream enters the chamber through a set of turning vanes that give it a whirling motion (i.e., like a cyclone), the flame has angular momentum, or swirl. Some burners provide only axial momentum; others provide swirl in varying degrees. Swirl can be obtained by delivering air into the annular channel around the fuel delivery tube through a tangential scroll; it can be obtained with hinged vanes in the entering air register (sometimes the vanes are adjustable); or the annular passage may have helical or screw-shaped inserts. Some burners have two concentric annular passages, one of which will be arranged with no swirl, the other with; there will then usually be a means of adjusting the fraction of air passing to each.

Axial Jet Flame

The axial jet flame produces a long slowly expanding, conical flame brush. It's envelope is bounded by the ability to mix the fuel and air to the required near-stoichiometric ratio, which is further discussed in Section 14.3 and 14.4 of the next chapter. For an unconfined flame, with a straight gas nozzle and a surrounding annular air tube having no swirl, with an excess air ratio of 30%, a flame length of about 5-1/2 feet will result when firing at a heat release rate of 1 MM Btu/hr. The width at the far end of the flame is about 1/3 that of its length. The flame dimensions can be scaled as the cube root of the heat release rate. Very approximately, the flame length,

$$L \simeq 5.5 \times \sqrt[3]{\text{Heat Release (in Btu/hr} \times 10^{-6}}$$

The flame is irregular and of ill-defined shape.

The flame length depends on the available oxygen supply -- the flame becomes longer as the excess air ratio is reduced to stoichiometric. However, the afterburner usually uses an excess of air, and only if an outside-air-burner is fired with low excess air will this effect be noticed.

When the free expansion of this jet is limited by the walls of the furnace enclosure, i.e., when the flame tends to impringe on the walls, the length of the flame is increased as compared with a flame firing into a large open chamber (see Reference 14-3).

The use of oil fuel can alter the flame pattern somewhat. With Number 2 fuel oil and fine atomization, the pattern will not differ greatly from that of the gas fuel burner. Most spray nozzles produce a conical spray pattern, which imparts a radial velocity to the jet; the result will be to spread the flame wider. Moreover, the conical spray tends to produce a low pressure zone immediately in front of it; with low velocity air and high pressure atomization, there will probably be a zone of recirculation at this point, that will tend to widen the flame spread even further. (See Figure 13-8.) This recirculatory pattern improves the stability of the flame, as mentioned earlier.

Reradiation by enclosing and confining the straight jet flame in a cylindrical box of hot refractory will increase the burning rate, so that less flame volume will be required; the required flame length in a box such that the flame just touches the walls would be somewhat reduced. Cold walls will decrease the burning rate and lengthen the flame and can result in a high residual concentration of CO, aldehydes, and unburned fuel in the combustion products, especially if the flame impinges on the cold wall.

Jet With Swirl

The long flame of the axial jet makes it awkward to use in the afterburner, and burners with swirl are commonly employed to reduce the flame length and the size of the chamber into which it fires. The addition of swirl can make striking changes to the flame shape and burning rates, especially if the annular entrance opening is divided into concentric zones (i.e., "primary" and "secondary" air) with separate control on the degree of swirl in each. Figures 13-8, 13-10a, b, and c show several markedly different patterns obtainable with swirl. Of these patterns only those of Figures 13-8 and 13-10 a are likely to be of use to the afterburner. Neither the Coanda nor the pencil flame burner is likely to contribute to good mixing conditions or the desired plug flow for the retention zone.

Swirl effects can be characterized by a swirl number, S, which compares the anguler momentum of the entering stream with the axial mementum (see Reference 18-5). For low to moderate values of swirl ($.1 < S < .5$) the flame develops a small internal recirculating pocket opposite the burner, as in Figure 13-8. Externally, the swirling flame narrows down beyond this pocket with a more crisply defined boundary. Its length decreases somewhat, especially with increasing swirl ($S \sim .5$), and its shape, becomes more like that of a cylinder, with diameter about 1/4 that of the length. (This flame has high stability, since the recirculating zone opposite the mouth of the burner returns hot flaming products containing highly reactive free radicals back to the starting point of the flame, thus helping to ignite the entering mixture.) As the amount of swirl increases still further ($S > 1$), the recirculating pocket enlarges, and the visible flame opens out into more of a cup shape, as in Figure 13-10a. The length will become less than half that of the axial jet, and the diameter will be roughly equal to the length. (This flame is relatively less stable since the central recirculation now returns relatively cool furnace gases rather than hot parts of the downstream flame containing the reactive free radicals.)

The transition in flame shapes becomes more complicated when the air (or fume) supply is subdivided into primary and secondary streams, each with its own swirl-generating register. When primary and secondary air have opposite rotation, the momentum of the combined stream becomes the algebraic sum of a + and - rotation. Greater turbulence can be generated, shortening the flame without forming the open cup pattern. Complex double-vortex ring recirculation patterns can be formed with different register settings. The pattern of the flame and its recirculation can also be affected by the shape of the surrounding ceramic burner tile or "quarl".

Interaction with Surrounding Enclosure

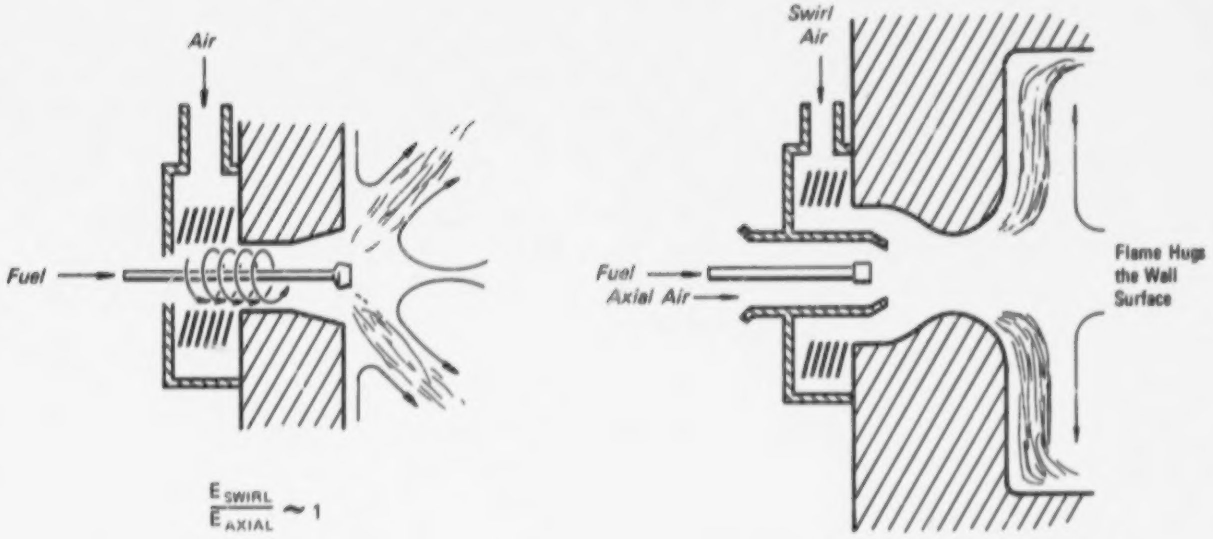
The forgoing discussion was premised on having an enclosure that is considerably larger than the envelope of the flame. An adjacent wall, or a closely confining chamber may simply constrict the flame, which impinges on the wall, with effects noted below. For flames that depend on gross recirculation effects to determine their shape, the wall effect can be more drastic, either reducing or completely suppressing the recirculation effect, and thus materially changing the shape and character of the flame.

In the event of flame impingement on the wall of the enclosure, if the wall surface is a hot refractory, and if the gross circulatory patterns of the firebox are not upset, the effect is not large. With a highly radiant flame the burning intensity may be increased, at the expense of decreased refractory life (unless the refractory material is specially selected to have the necessary temperature resistance; because of the turbulent nature of such a flame, the surface is alternately swept by hot licks of flame and cold eddies of unburned gas; this can cause a thermal fatigue effect, causing the refractory surface to spall away). Castable refractory materials can also be locally overheated to the point of de-bonding the cement, causing spalling.

In the event of flame impingement on a cold surface, as in a metal enclosure that is cooled by sweeping the incoming fume stream over the back side of the wall, the hot spots can cause severe warping of the metal, and possibly cracking it from thermal fatigue. Additionally, the cooling effect of the wall quenches the flame before the combustion reaction is complete, leading to the formation of carbon monoxide, aldehydes, etc., which must then subsequently be destroyed in the holding section of the afterburner if the unit is to satisfy its anti-pollution requirements. The cold surface also acts as a trap, or sink, for the deactivation of the free radicals, which serve an important function in the formation of the flame, and may also play an important role in the destruction of the fume.

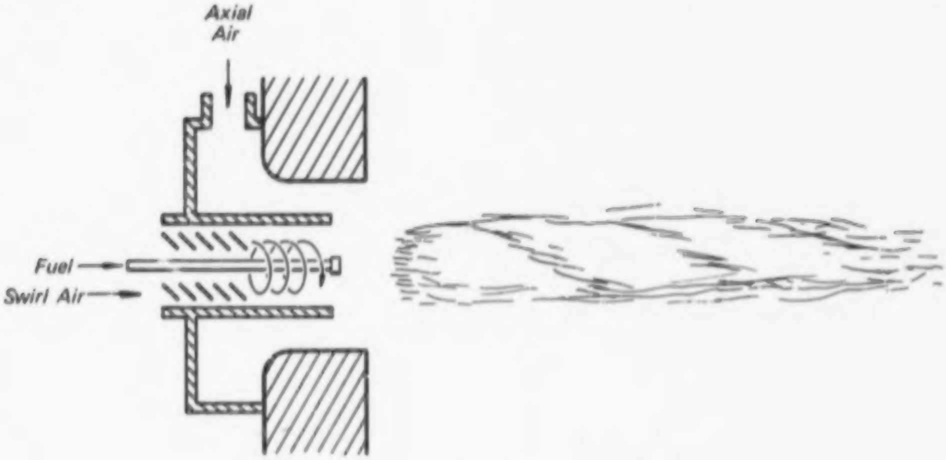
A large afterburner equipped for oil firing may require more than one burner; as the number increases, the array becomes a distributed-source arrangement. An array of two, three, or four, still behave as adjoining discrete burners. Placed side-by-side, their plumes will interfere--especially if the flame pattern is one that involves a considerable degree of recirculation. With flames having no swirl, and with little or no recirculation, the individual plumes are merely squeezed and narrowed. Swirling flames, when all flames have swirl in the same direction, tend to join to form a single large vorticular pattern; hence (assuming this is undesirable), the adjacent burners should be arranged for swirl in alternate rotational directions. The effect of gross rotating flow patterns in the combustion chamber and retention space can have important consequences in the mixing-in of fume material into the flame, and in the degree to which one can obtain a plug-flow condition in the retention space; this is discussed in later sections.

Figure 13-10 follows


Air
Fuel
$\frac{E_{SWIRL}}{E_{AXIAL}} \sim 1$
Swirl Air
Fuel
Axial Air
Flame Hugs the Wall Surface

A. **Air Entry with Swirl** (Conical Flame)

B. **Air with Swirl on Outer Boundary** (Coanda Burner)


Axial Air
Fuel
Swirl Air

C. **Air with Axial Jet, Swirl in Center**

Figure 13-10. EFFECTS OF SWIRL

Burners are sometimes arranged to enter the combustion chamber from the side; this is done in vertical "stack" units; the fume passes straight through, while the flame passes across it. This presents problems especially from the point of view of the mixing the flame and the fume streams. The flame possesses considerable momentum in the direction of its axis, and the wall constraint effect is substantial, as shown in Figure 13-11.

13.3.3 The Distributed Burner

The distributed burner is simply a multiple array of discrete burners. The flame pattern in each burner of the array will reflect the pattern of its discrete element, except that it will be altered by being placed close to its neighbors. Distributed burners are designed only for gas fuels. Some individual versions were described earlier in the section on flame stability; the Maxon "Combustifume", the Hirt "Multi-jet" burner, the North American "Flamegrid", and the Hydro-Combustion Corporation gas burners are sketched in Figure 13-12. For all practical purposes the flame can be considered as extending no more than a few inches from the open end of the burner structure. At low firing rates, the flames are largely confined within the burner structure unless the fume flow is reduced to the point where slow mixing occurs. Then a long "lazy" flame may result.

13.3.4 Discrete Burner With Precombustor

With increasing limitations on burning gas fuels, oil fuels may become essential. On very large units there may be occasion to consider the use of residual-oil fuels. Residual oils are much harder to burn without generating smoke and forming high NO_x and SO^2. There is limited data on the NO_x formed in afterburners. Satisfactory results in avoiding the smoke have been obtained using oil burners in conjunction with a small refractory-lined "precombustion" chamber. The burner head delivers the air and atomized fuel into this chamber with a fairly high velocity (as much as 25 inches of water draft may be required) and with a carefully pre-set degree of swirl. Combustion occurs largely within the chamber, under conditions of high turbulence (in some designs generated by counter-rotating vortices), and with high temperature re-radiation from the very hot refractory walls. The flame does project from the chamber opening, but it extends only about one or two opening diameters. The energy release rate within the chamber is very high, of the order of 10^6 to 10^8 Btu/(hr-ft^3), and the high temperature of the refractory requires special materials. Burners of this general type are supplied by Thermal Research, and John Zink.

13.3.5 Scale Factors and Energy Release Rates

Burner manufacturers specify the fuel rates that can be handled in their units. Burners are usually supplied in standard "frame size" units, and one can adapt these to a range of conditions by changing the size of the gas flow orifices, nozzle tips, and air register settings, etc. In the case of the distributed burner, ratings are given for a given burner box element, which can be incorporated in multiple arrays as needed for increased capacity. To specify the capacity, the burner manufacturer must know the fuel to be used, the air or fume stream (if used in the burner) temperature, the nature of the fuel components and their heating value that may be carried in the fume stream. He must also know the pressure drop or draft available on both air/fume and fuel streams, or he may specify requirements for his burner, which must be provided in the installation. The variety of burner designs and their pressure requirements is so large that no general rules can be given here.

Typical heat release rates for discrete burners range from a few hundred thousand to several million Btu/hr, both in oil and gas fired units. For the distributed burner, Maxon lists 500,000 Btu/hr per lineal foot of the burner assembly; other makes are equivalent. Although the burner manufacturers often work on the basis of a heat release rate per unit volume of the combustion chamber, such numbers are not relevant here, since the chamber volume is governed by the space required for mixing of flame and fume components, and for holding them at elevated temperature for the length of time that is necessary for the destruction of the pollutants. For scaling purposes, it is sufficient to recognize that the capacities of the burner and the afterburner unit will vary with the volume, or the cube of the linear dimension, for units which have dimensional similarity.

13.4 Installation

13.4.1 Setting

A great variety of arrangements are used by the different manufacturers for mounting and supporting their burners and for introducing the flame into the combustion chamber. The manufacturer's installation drawings should be referred to.

13.4.2 Piping and Controls

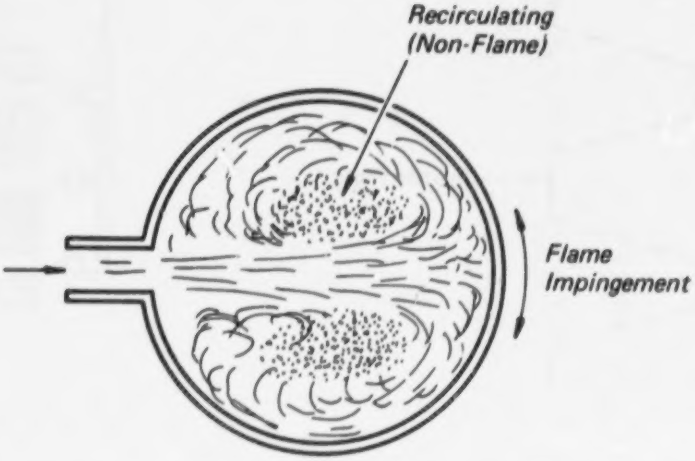
Refer to Chapter 9 for a review of requirements that must be met for control and safety. Fuel and combustion air, control valves, shutoff values, pilot, and flameout detector are included in some burner assemblies as package units. Alternatively, the afterburner manufacturer may assemble his own version of these components. Whatever the arrangement, the purchaser carries the responsibility of installing a safe operating unit.

13.4.3 Initial Operation

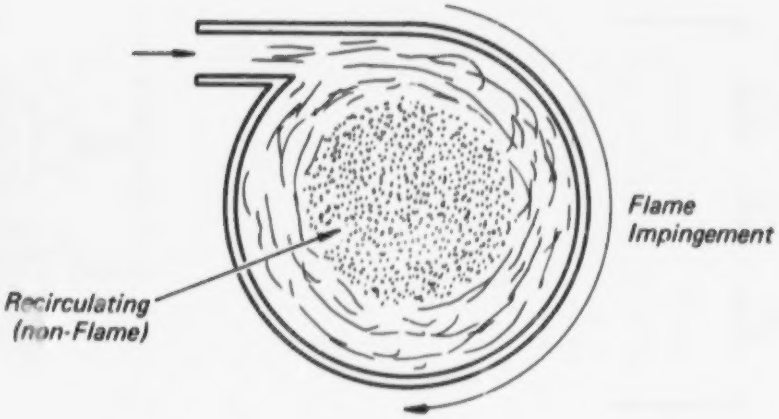
Afterburner manufacturers regularly include supervision by an experienced operator for the initial startup for a stated time interval as part of their contract. Additional services are available for a fee. On larger and more complex units, the purchaser should also obtain a manual of standard operating procedure. Maintenance instructions, and service parts lists are also necessary items.

The initial operation of the burner on some units will be a simple operation of pushing a button, and making trim adjustments on the air/fume damper settings. Other units may be more complicated, involving a schedule of slow initial heating to cure the cement used in refractory linings. All emergency alarms and shutdowns should be proven by test.

Figures 13-11 and 13-12 follow

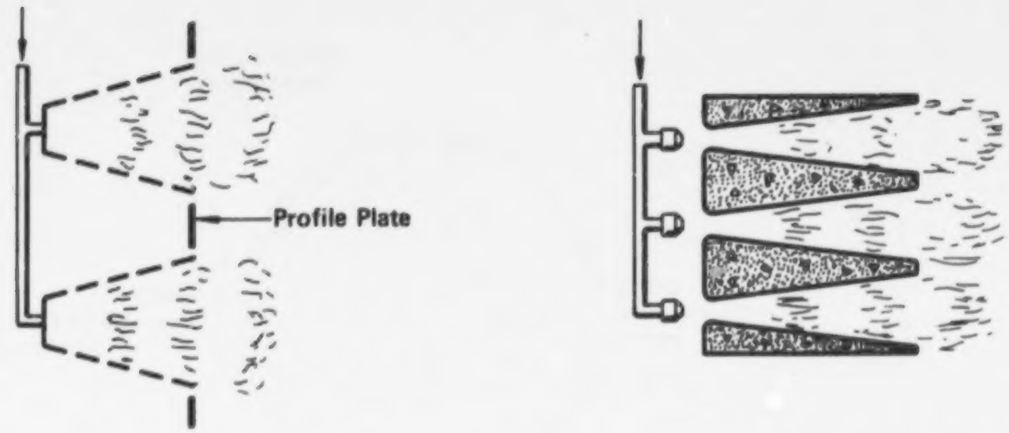

Recirculating
(Non-Flame)
Flame
Impingement

A. Radial


Flame
Impingement
Recirculating
(non-Flame)

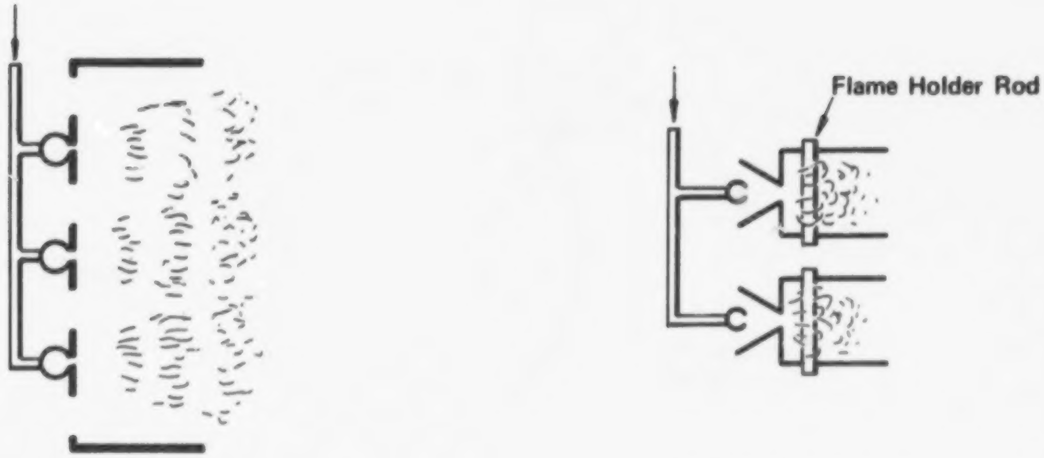
B. Tangential

Figure 13-11. FLAME PATTERNS - SIDE ENTRY


Profile Plate

A. Maxon "Combustifume"

B. Hirt "Multijet"
Gordon & Piatt "Turbofire"


Flame Holder Rod

C. North American "Flame Grid"

D. Hydro Combustion

Figure 13-12. DISTRIBUTED FLAME GAS BURNERS

Once initial operation has been completed, verification tests on emission control by the afterburner are desirable, and may be required by the pollution control agencies. Some adjustments of the burner controls may be necessary to achieve the required emission control at this time. It is very desirable to have this breaking-in and adjustment completed before the control officer is asked to approve the installation.

Chapter 20 of this manual gives test procedures for determining the emission control performance of the afterburner.

Chapter 14. MIXING OF FLAME AND FUME

14.1 Introduction

In the preceding chapter it was noted that the afterburner process, for convenience of analysis, was broken into flame, mixing, and retention processes. This section discusses the second, or mixing process. It bears to be repeated that in the actual afterburner these three processes are not separate and distinct. Particularly, as regards the mixing operation, the formation of a flame and the mixing-in of the fume may be closely tied together, especially when the fume stream is used as the source of oxygen for the combustion. Also to be repeated is the role the flame may play in initiating the chemical breakdown of the fume constituents because of the presence of the free radicals in the flame; this role is minimized in the present discussion (see Section 12.2.3). It is assumed that uniform mixing of combustion products and of the remaining bypassed fume stream is essential for proper performance of the afterburner. Anything less than this results in a possible failure of the unit to control the pollutant emissions, and a wastage of fuel since the temperature setting of the afterburner stack must be set excessively high.

Also repeated: The entire fume stream, even if used as the oxygen source, cannot be passed through the flame, since the mixture ratio of fuel and fume lie outside the flammable limits for temperatures in the range used in the afterburner (see Section 12.1.3). Hence, at least a portion of the fume stream must be passed around the flame and blended in subsequently. (In some "excess air" burner designs, the excess air or fume is bypassed through a channel within the burner unit, however.)

14.2 Characterizing Mixedness

An important and difficult question faces the designer who is attempting to insure good mixing in his afterburner. What constitutes good mixing? In terms of the concentration of the flame products to be mixed into the fume, does he want to achieve a uniform blend with a maximum deviation of 10%, 5%, 1%, or 0.01%? The answer must depend upon the objective of the mixing. If the mixture leaves the mixing section of the combustion chamber with a non-uniformity in temperature amounting to 100°F, the chances are that this disparity will remain and carry part way through the holding section; and if 1400°F is required for the time available to destroy components of the fume stream, then the temperature setting on the controller would have to be set at 1500°F to attain the necessary destruction. If, on the other hand, the fume stream contains a highly toxic vapor that must be controlled to the parts-per-billion level, mixing deviations at the 0.01% level may be needed.

By way of definition, concentration here refers to the fraction of the material in one stream that is being blended with a second, and which remains in a unit volume (or mass) of the mixture at any point in time (or position) after the mixing has begun. If one arbitrarily assigns a value C = 1 to the first stream, C = 0 to the second, then a number between 1 and 0 will characterize the

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degree of mixing achieved at any downstream point. The local mixing effectiveness can be defined as 100 $(1-C)/(1-C_{mixed})$ in percentage points, for mixing the first stream into the second. Clearly this definition will not apply upstream from the mixing. Referring to the question posed above, it is necessary to determine if the mixing effectiveness required is 90%, 95%, 99.99%. Unfortunately, the mathematical models and experimental correlations cannot be trusted with anything like the predictions for 99.99%. The following discussion will attempt to make it clear why this is so.

Mixing between two fluids is actually a process of subdividing and recombining on a random basis. As the subdivisions become smaller and smaller the fluids will finally become mixed on a molecular scale. For chemical reaction between fume stream components and oxygen, complete molecular-scale mixing will ultimately be required if pollutant components are to be fully destroyed. We may start this by mechanically subdividing the streams (i.e., the distributed burner--we can only go a little way with the distributed burner; however, it is an important way in terms of saving time and space). We then depend on natural hydraulic blending through the mechanism of eddies, and finally depend on molecular diffusion.

Eddies are relied upon for the predominant part of the mixing process. Eddies are generated between parallel moving streams when the streams have different velocities. The velocity gradient between the streams generates shear forces, which in turn cause a circulating motion that becomes an eddy. The eddy tends to carry material with it across the line of flow. Naturally, high velocity fluid from one stream has momentum which thereby becomes transferred over into the adjacent low velocity stream. Similarly, low velocity fluid acquires momentum as it is carried over into the high velocity stream. Both material and momentum are thereby exchanged between the two streams, with what are initially rather gross eddying motions. Continued motion of these eddies result in their breakup or decay into a series of smaller and smaller eddies, with finer and finer distribution of material. The process as a whole is a form of turbulence, and the scale of the turbulence is measured by the size of the eddies that predominate at a point. If one can pick out a streamline which describes the smoothed, average flow path of a portion of the stream, unaffected by the random motions imposed by the eddies, one can imagine a ball riding and floating in the fluid, and following only the streamline. The ball would then see a localized movement of fluid past it, whose velocity and direction would depend on the eddying motion in its neighborhood. The time averaged value of the square of these unsteady velocities would be a measure of the turbulence intensity (it will be seen that this intensity is a measure of the kinetic energy associated with the eddying motion). For a more detailed review of this complex subject, see Reference 14-1).

The mobility of an individual molecule with respect to its neighbors, under the impetus of its own velocity, is relatively low; it keeps colliding with the many other molecules nearby. That is, the rate of molecular diffusion is low, especially compared with the gross eddying motion we have described above. Diffusion becomes important primarily after the eddies have broken up to a very fine scale, and the local turbulence intensity has diminished greatly. The chemical reaction that is desired from the afterburner can be completed only after the final diffusion mixing process is largely completed for it is only by diffusion that the oxygen and pollutant molecules can find each other.

The analytical models and experimental correlations describing the mixing process quantitatively give only the time-averaged value of the velocity, concentration, and temperature in rather specialized cases. Experimental measurements made on tests of afterburners are usually obtained with thermometers, pressure probes, and sampling devices that cannot follow the rapid fluctuations of the eddying motion. Highly sophisticated laboratory instruments have been used to follow the turbulent velocity and temperature fluctuations for frequencies up to several thousand cycles a second, but these have not been used to study afterburner mixing. No instrument can tell the effectiveness of mixing on the molecular scale, which is what ultimately counts. Thus, small eddy-pockets can be formed that will persist in composition and temperature for some distance beyond where the correlation has predicted uniformity. And mixing on a molecular scale (and uniformity of temperature on a micro-scale), may not be achieved for some real distance beyond this. Also, real combustion chamber shapes deviate from the idealized model, and corners and bends can introduce gross changes in the flow pattern; on a smaller scale, wall roughness and boundary-layer effects are not accounted for.

In a paragraph above we gave a criterion for measuring the degree of local mixing (C a number between 0 and 1). This does not apply for the mixing as a whole. There is no simple, rigorous, easily calculable criterion. Using the deviations of concentration or temperature from the long-term mixed values, or from experimental values found in a traverse, can one find an average (if so, how weighted?), or a mean deviation? Refer to Figure 14-1 for three different temperature profiles that might be obtained across an afterburner chamber. Profile (a) might be found in a fairly well mixed afterburner, showing a temperature peak opposite a burner. Profile (b) shows a nearly uniform, but slightly skewed pattern. Profile (c) shows cool, unmixed zones along the walls. Using a weighted average temperature (weighted according to cross section area--that is using

$$T_{avg} = \frac{\Sigma\ T\ (\text{area element})}{\text{whole area}}$$

where the cross section has been divided up into equal area elements), one can find a maximum deviation from the average, d_{max}. We suggest using a ratio,

$$100\ (\frac{d_{max}}{T_{avg}})$$

for determining the mixing effectiveness of the chamber as a whole (or a similar ratio based on concentration). By this criterion, curve (b) would be the best choice for the afterburner; however, common sense would dictate that (a) would be the choice (the small peak will do no harm). Clearly profile (c) is the least satisfactory.

To initiate a high rate of mixing, one should generate the large scale random eddy patterns in the stream which generate the turbulence. This is brought about by large velocity differences in the fluids to be mixed. The high velocity means high kinetic energy--which makes another point. Mixing costs energy, which is supplied by the fan or the pressure of the flowing gas. More rapid and more effective mixing will require more energy. A casual examination of the equations of Figure 14-2 suggest that the mixing effects are only a function of the distance from the nozzle; further thought recognizes that for a given mass flow rate the higher velocity uses a smaller jet diameter, thereby reducing the size of the entire operation. Since the jet velocity, V_o, varies inversely with the nozzle area, i.e., as $1/D^2$, the distance, L, require to achieve a given degree of mixing will vary as 1/ V_o, and inversely with the rate at which energy is supplied.

As a general observation, it can be said that the process of mixing flame and fume streams to the point of having a uniform temperature for decomposing the pollutants is the most difficult part in the design of the afterburner. It is the least well understood and the least effectively handled in many working units. Its accomplishment is rarely demonstrated by adequate tests in afterburner units; its inadequacies show up as residual effluents in the stack and wastefully high temperature settings.

14.3 Axial Jet Mixing

In order to understand mixing behavior of various afterburner designs it is helpful to work with idealized models. Although these models may be unrealistic in describing in detail the complex behavior of the many burner configurations, they give an insight as to what to expect, and the effects of variations in arrangement can be examined for the ways in which they alter the general pattern of behavior exhibited by the idealized models.

The first model is shown in Figure 14-2. It consists of a single "free" axial jet stream without swirl discharging from a cylindrical pipe into an open space filled with fluid at rest. The velocity distribution at various distances away from the nozzle are shown on the sketch (see Reference 14-2). Immediately downstream from the nozzle exit there is a "potential core" of moving fluid in which the velocity remains undiminished. The core is eroded by the eddy mixing process, but does not fully disappear for a distance of about 6 nozzle diameters downstream. Thereafter, the centerline velocity decays logarithmically. Meanwhile the eddies spread the jetted material laterally, and the zone of mixed-in material increases in width. The concentration of mixed-in material is greatest along the centerline of the jet, diminishing to the vanishing point at the edge of the jet stream.

The model describes with good accuracy the distribution of velocity, temperature, and concentration (temperature and concentration spread a little faster than the momentum, as measured by the velocity). The distribution shown represents a time averaged one and the corresponding time-averaged streamline is a straight line. Because of the eddy motion of the mixing process, the actual trajectory of an individual particle only approximately follows the streamline; it wanders and weaves in an irregular path as shown in the figure.

This model is helpful because it shows what happens when a discrete burner discharges its flame into an open combustion chamber without swirl. The jet here represents the flame plume, which does not necessarily have the same diameter as the opening of the burner (the effect is close, however, if the jet

Figures 14-1 and 14-2 follow

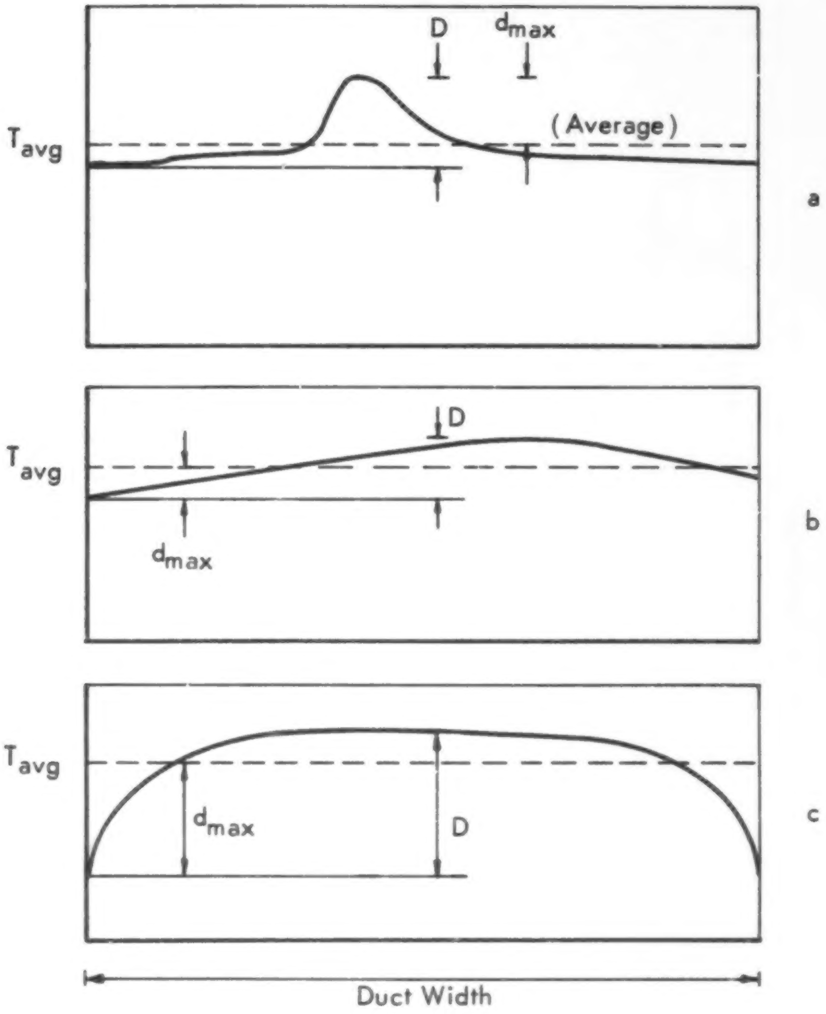
D
d_{max}
T_{avg}
(Average)
a
D
T_{avg}
d_{max}
b
T_{avg}
d_{max}
D
c
Duct Width

Figure 14-1.

228 b

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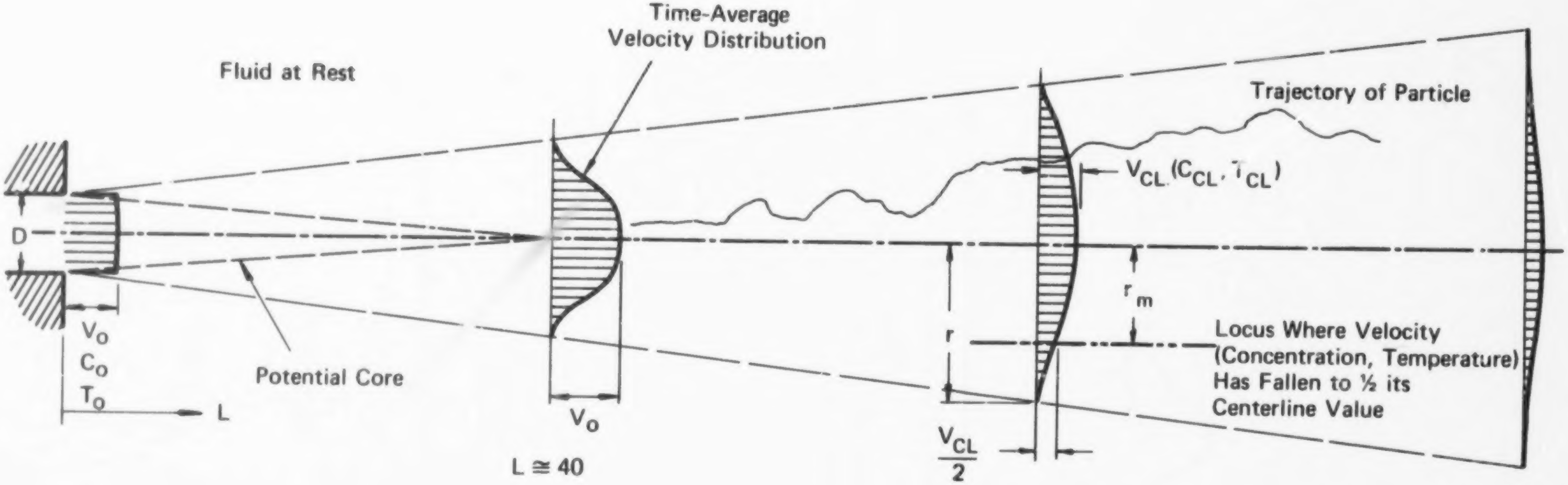
S-14121
67784

228C


Time-Average
Velocity Distribution
Fluid at Rest
Trajectory of Particle
D
V_{CL} (C_{CL}, T_{CL})
V_o
C_o
T_o
L
Potential Core
V_o
L ≅ 40
r_m
r
Locus Where Velocity
(Concentration, Temperature)
Has Fallen to ½ its
Centerline Value
$\frac{V_{CL}}{2}$

Beyond the Potential Core Range, at Section A:

$$\frac{V}{V_{CL}} \approx \frac{C}{C_{CL}} \approx \frac{T}{T_{CL}} = \frac{1}{2}\left(1 + \cos\frac{\pi r}{2r_m}\right)$$ .... *Lateral Profile Cross Section*

$$r_m = L/8$$ .... *Lateral Spreading Rate*

$$\frac{V_{CL}}{V_o} \approx \frac{C_{CL}}{C_o} \approx \frac{T_{CL}}{T_o} = \frac{4D}{L}$$ .... *Decay of Centerline Velocity*

Figure 14-2. MIXING OF A FREE JET IN AN OPEN CHAMBER

velocity and size is corrected for the high temperature of the gas (Reference 14-3). Clearly apparent is the very large distance required to dissipate the jet velocity and mix the flame into the contents of the firebox. Roughly 20 times the flame diameter is required of this mixing to reduce the core concentration to 20% of its initial value.

14.4 Concentric Axial Jets

The second idealized model is that of two concentric jets, as given in Figure 14-3. There is a central core jet, entering with a higher velocity, an inactive gap, and an annular jet surrounding it, entering at a lower velocity. The jets have symmetry about the centerline of the core jet. There is no swirl, and the jets discharge into a large, stagnant chamber.

This model shows that a relatively stagnant recirculating zone develops in the gap between the two jets (Reference 14-4). This gives a low pressure area that diverts the two jet streams, so that beyond b they join to form a single jet. Thus, the second model approximates the behavior of the single jet model when sufficiently far away from the nozzle exit.

The concentric jet model represents the situation of a flame (i.e., the inner jet) mixing with a secondary air or fume stream (the outer, annular jet). The velocity profiles of Figure 14-3 are insufficient to describe the mixing between the two jet streams; Figure 14-4 shows that the velocity profile at section a-a is actually made up of two profiles, a'-a' and a"-a", from which an idea can be gained of the degree of interpenetration that has occurred. A temperature profile of a hot core stream mixing with a cold fume stream would look like the profile a'-a'. As with the single jet model, a distance of at least 20 diameters of the core jet (the flame) is required to get an effective blending of the two streams.

To account for the effects of putting the burner in an enclosure, consider a third model (refer to Figure 13-8). Jets are introduced into a combustion chamber of finite dimensions. No longer is there an infinite sink into which to dissipate the momentum of the jets. Material from the surroundings entrained into the jet stream by the eddy mixing process must be replaced; it comes from developing a toroidal recirculating pattern in the corner of the chamber, around the jet openings, so that there is a recycling of material taken off the outer envelope of the jet stream. Far down the chamber the concentration and temperature approach a homogeneous mixture of the two streams.

The presence of the enclosure makes it hard to predict the effects on mixing. Data is available for the mixing of two concentric streams mixing in a pipe (Reference 14-5). Decay of velocity and concentration of the center stream (downstream from the nozzle) can be calculated, but the procedure is involved.

14.5 Subdivided Parallel Jets

A multiple burner installation may be simulated with a model consisting of an array of confined coaxial jets of the type discussed in the preceding section.

The simulation can only be qualitative, since no array can give a true radial symmetry for each element of the array. It also suffers in that with the single coaxial jet model, the presence of a wall contributes considerable turbulence from its wall boundary layer, which is not present when placing the jets side by side. The single unconfined jet has a large reservoir of dead air surrounding it with which to exchange momentum that will slow it down; adjacent moving streams can provide no such braking action. Hence, parallel arrays of jets are not as effective in promoting mixing as one would predict if scaling from a single large jet to determine the performance of an individual small jet in the array. In one case (Reference 14-6), the use of 64 small nozzles mixing gas into a secondary stream in a pipe was found to decrease the distance required to obtain a mixed condition by a factor of 5 as compared with a single nozzle mixer; scaled from the free jet model, one would have anticipated an 8 fold reduction. This particular test was based on very low jet velocities, however, so that pipe turbulence rather than jet turbulence controlled the mixing; the same numbers would not necessarily apply to the multiple burner case. For an accurate determination of mixing, the designer should make model tests of his own configuration, and verify the final design with carefully made traverses on the actual afterburner.

Despite the difficulties in predicting the degree of mixing obtainable with an array of parallel jets, the principle is one of the most useful for obtaining good mixing where space is limited, i.e., the multiple burner approach should be used wherever possible.

A precautionary note: Some afterburner units are designed with multiple jet type gas burners that form an array of the general type discussed. If the entire fume stream is passed through the individual burners or between them, the principles discussed above apply. If, on the other hand, the fume stream passes in part around the outside of the burner array assembly, then the mixing must be considered from the standpoint of a single large burner source.

14.6 Profile Plates

Profile plates are used to control the bypassing of fume stream around distributed gas burners when the whole stream cannot be passed through the burner (see Figure 14-5). Burner exit stream and fume stream entering by way of the slit between the burner box and the profile plate can be considered as parallel rectilinear jets. The problem here is the distribution of the bypassed material into that issuing from the burner opening. Approximately 20 times the width of the slit opening should be allowed for nearly complete uniformity. The effective opening width is less than the actual one because of the sharp-edged-orifice effect; also the turbulent condition of the flame leaving the box, tends to reduce the number of L/D's required for this mixing process, but these effects should not be counted on. Test data obtained by Maxon on their "Combustifume" unit show a nearly uniform temperature distribution on a section taken 42 inches away from the profile plate (temperature variation 990°F to 1007°F); at a distance of 18 inches, a temperature varation of 998°F to 1140°F was measured across the channel. Assuming a slot width of 1-1/2 inches and an effective jet width of 0.9", their results would be consistent with the recommendation just given.

Figures 14-3 through 14-5 follow

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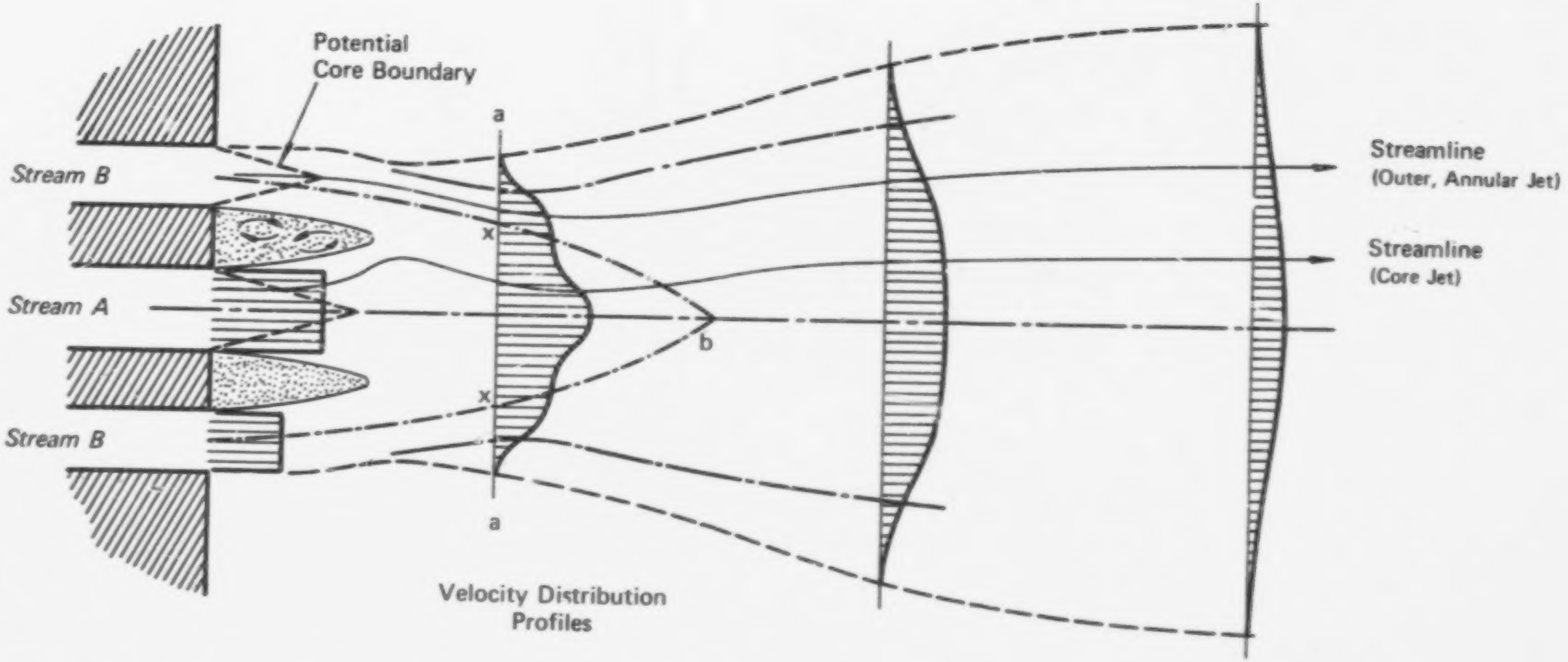
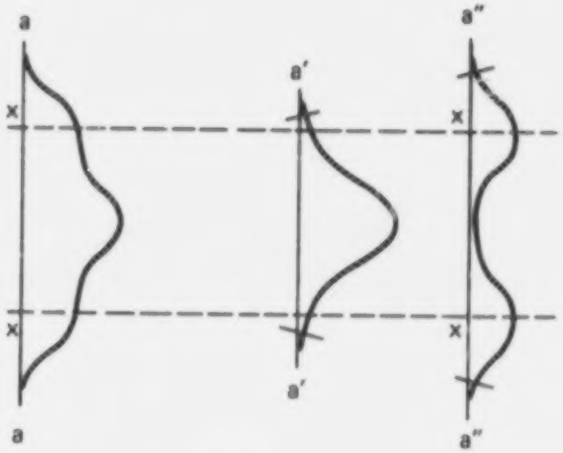

Potential
Core Boundary
a
Stream B
x
Stream A
b
x
Stream B
a
Velocity Distribution
Profiles
Streamline
(Outer, Annular Jet)
Streamline
(Core Jet)

Figure 14-3. MIXING OF CONCENTRIX AXIAL JETS (REF. 14-4)

230½



Velocity Distribution
at Section a-a
(See Fig. 14-3)

The Velocity Distribution is the Sum of the Two Distributions from Inner and Outer Jets, A & B; Also a′-a′ Represents Concentration Profile of A, and a″-a″ that of B.

Figure 14-4. VELOCITY COMPONENTS IN CONCENTRIC AXIAL JETS

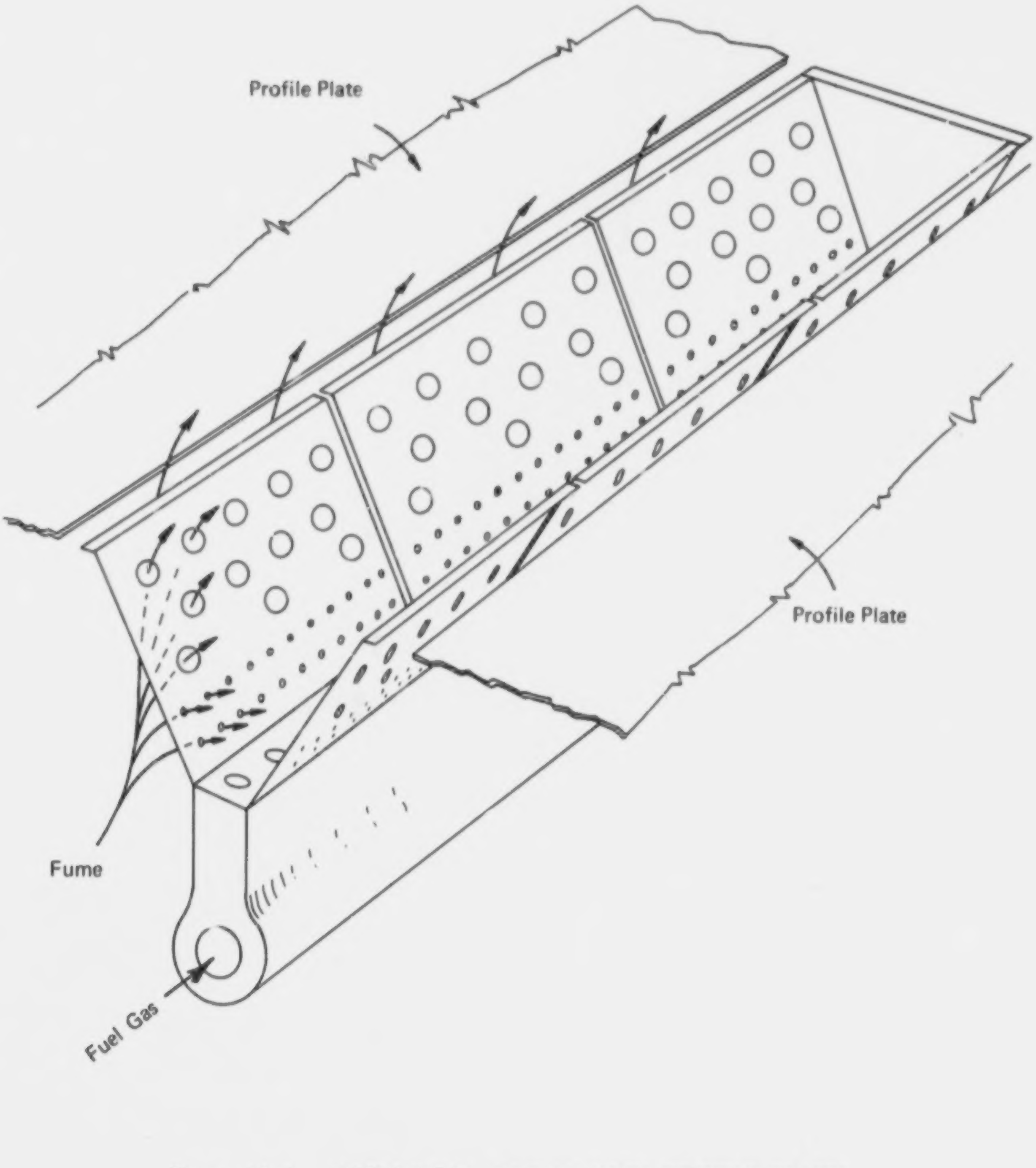

Profile Plate
Profile Plate
Fume
Fuel Gas

Figure 14-5. DISTRIBUTED BURNER WITH PROFILE PLATES

230 d

14.7 Cross Stream Jets

Another approach to improve mixing is to introduce cross-stream momentum into the system, to carry the one stream into the other by a mechanism more powerful than the eddies generated by velocity gradients between parallel streams. This can be done by directing the streams at an angle to each other, or it can be induced by baffles.

Figure 14-6 shows results of an experiment (Reference 14-7) measuring the rapidity with which a jet stream entering from a side wall will mix with a main channel stream. The jet stream is deflected into a curved path, but penetrates across the channel to a considerable depth. The degree of turbulence is greater than with the parallel jet, and the jet velocity decays much faster than for the parallel jet case. Also, downstream from the side-entering jet a wake pattern forms which includes a double line of counter-rotating vortices which aid in the cross stream mixing. Where the cross stream jet differs appreciably in density from that of the main stream, the dimensionless correlations of Reference 14-8 should be used. See also References 14-19 and 14-20.

The model tests indicate that when using a discrete burner unit, the fume stream would be better blended into the flame through a series of cross-stream jets, perhaps arranged from alternate sides as in Figure 14-7 so as to avoid the impinging effects shown in Figure 13-12). In rectangular array, this might be applied as in Figure 14-8, particularly as applied to a "line burner".

14.8 Baffling

Baffling can be used to induce cross-channel mixing; however, its use must be understood, otherwise it is costly in added pressure drop and in wasted chamber volume used up in recirculatory "dead zones". Its application should be made with full regard for the possibilities of impinging the flame on the baffle, leading to excessive maintenance problems.

Baffling must be designed to divert the stream laterally an amount equivalent to the lateral scale of unmixedness. That is, if the variation of concentration or temperature to be evened-out spreads from the center of the chamber to the wall, then the stream needs to be diverted by at least this amount. Checkerworks or grids are sometimes used with the object of improving the mixing; they are not very effective for this purpose, since they promote mixing on the scale of the grid of the checker or screen; they will not help much in exchanging material across the whole channel. On the other hand, grids and checkers are very effective in the small-scale mixing, and in improving the flow distribution and velocity distribution in the channel (see References 14-9,10).

The following baffles are commonly used.

14.8.1 Bridge Wall Baffle

A dam or bridge wall placed across the chamber is a fairly effective mixer. For best results, bridge walls should be used in pairs, as in Sketch A of Figure 14-9. Used singly, the bridge wall tends to acquire a recirculating zone on its back side, and the main stream may not mix well across the stream lines, as in Sketch b. With a single bridge wall partition having a blockage of 60 - 75%, good mixing was achieved in a distance of 10 times the duct height, according to Reference 14.9. The presence of a high velocity core flow ahead of the bridge wall, as from a burner jet, helps somewhat to break up this streamline flow pattern downstream from the dam, where a pair of counter-rotating vortices develop, as in Sketch c. The bridge wall should extend at least half way up the chamber and the second wall (if used), should be placed at least one chamber diameter (or width) downstream. For very high effectiveness mixing (99.9%?) the bridge walls should reach almost to the opposite sides of the chamber, giving a high velocity to the gases, and the distance between the two walls should be increased. The chamber formed between the two bridges then becomes a "mixing chamber", with highly turbulent flow generated by the dissipation of the kinetic energy of the flow through the narrow openings. Such an arrangement can be costly in terms of pressure drop and space required, but this may be justified if high effectiveness of mixing is required.

14.8.2 Ring and Disc Baffles

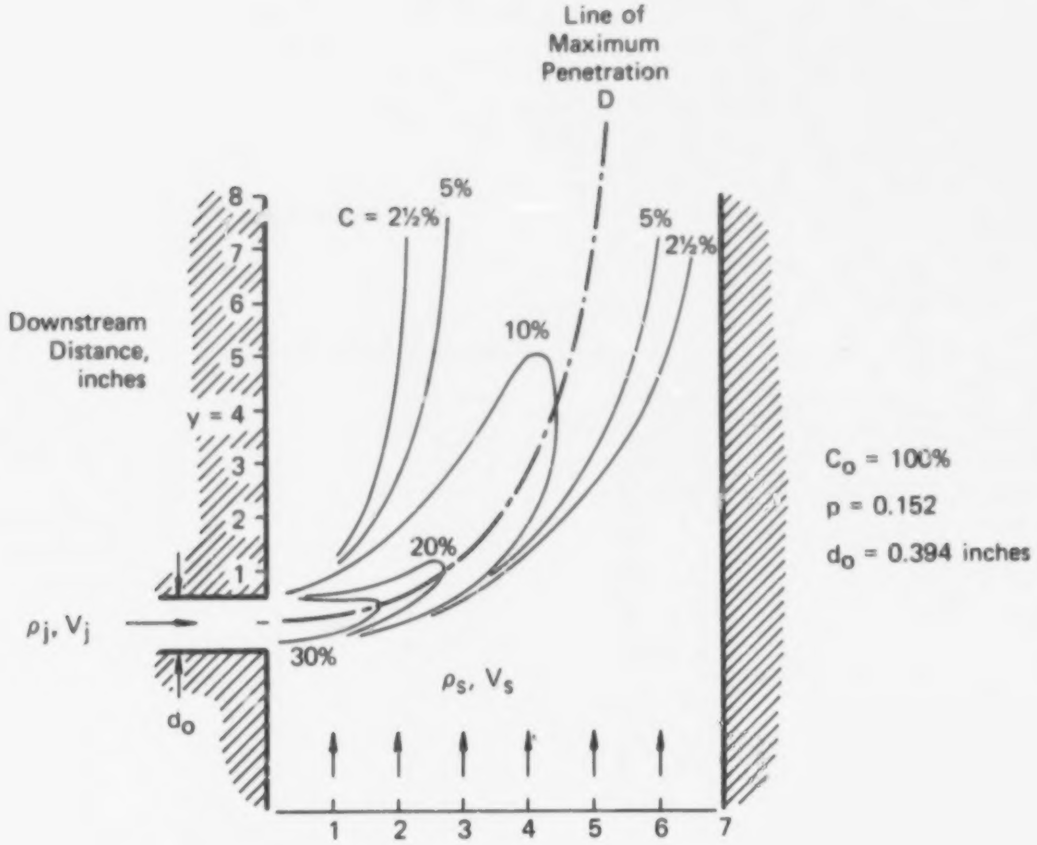
A series of rings and discs inserted in a pipe, as in Figure 14-10, has been in common use for blending and mixing in pipeline reactors. One or two sets of ring-and-disc should give effective mixing and act similar to pairs of bridge walls. It is suggested that the ring and disc each act as orifices and each should have a free-flow area about half that of the chamber for good mixing. Spacing should be no closer than half the minimum width of the chamber. As with flow through an orifice the pressure drop across such a mixer can be estimated by finding the velocity of the hot gases through an area which is 0.6 times the free-flow area (vena contracta); then, for a mixed gas density w, lb_m/ft^3, the pressure drop for each ring and disc pair should be about wV^2/g_c. Space required for mixing is about 10 chamber diameters for a ring and disc pair, which makes it costly in space.

A modified version of the ring and disc is used by one afterburner manufacturer, Figure 14-11. Ring and disc are aligned to form an annular orifice, which is placed above a ring burner, as in Sketch a. The arrangement is very similar to the idealized free jet model of Figure 14-2, except that an annular opening is used; furthermore, the full annular opening width cannot be used for the jet width dimension because the opening acts like a sharp-edged orifice. d should be taken at 0.6 times the annular gap width. The manufacturer's data on velocity, concentration, and temperature profiles (Reference 14-12) is given in Sketches b, c, and d of Figure 14-11. The values of L/d on the basis just described would be,

Distance	L/d
5 inches	2.14
12 inches	5.1
24 inches	10
60 inches	26

d is the effective gap dimension, allowing for vena contracta.

Figures 14-6 through 14-12 follow


Line of Maximum Penetration D
C = 2½%
5%
5%
2½%
10%
20%
30%
Downstream Distance, inches
y = 4
8
7
6
5
3
2
1
ρ_j, V_j
d_o
ρ_s, V_s
C_o = 100%
p = 0.152
d_o = 0.394 inches
1
2
3
4
5
6
7

For the Lines of "Maximum Penetration of Concentration", OD,

$$\frac{y}{d_o} = p^{-0.85}\left(\frac{x}{d_o}\right)^{0.34}$$

WHERE $p = V_s / V_j$

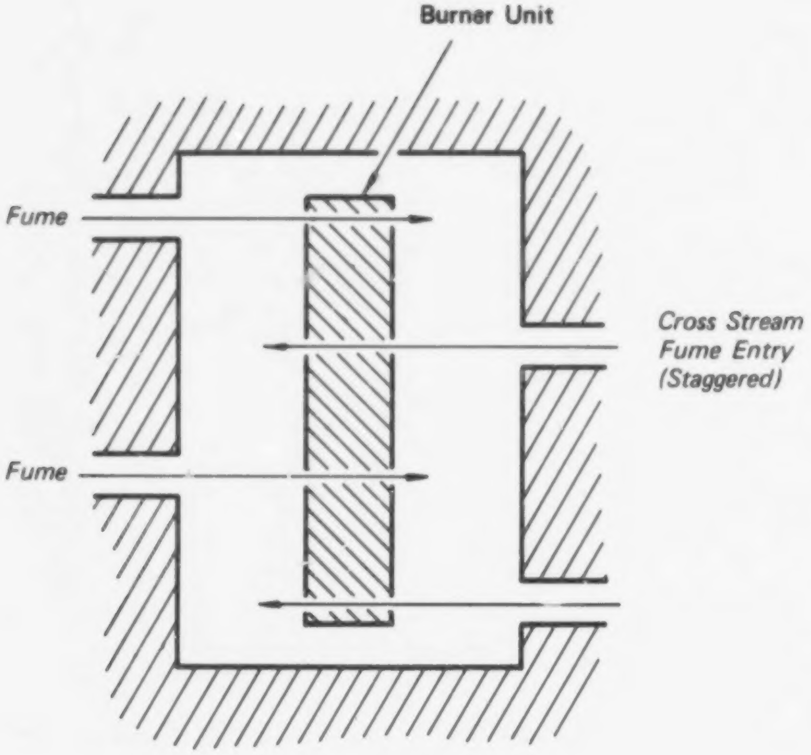
V is Velocity

ρ is Density

REFERENCE:

M.A. Patrick; *"Experimental Inv. of Mixing and Flow....."*, *Journal Inst, Fuel*, v. 40, Sept. 1967, pp. 425-432

Figure 14-6. SIDE ENTRY OF A JET INTO A MAIN CHANNEL


Burner Unit
Fume
Cross Stream
Fume Entry
(Staggered)
Fume

(Section of Box – Looking Toward Flame)

Figure 14-7. SUGGESTED ARRANGEMENTS FOR CROSS-STREAM MIXING OF FLAME AND FUME IN RECTANGULAR BOX

232C

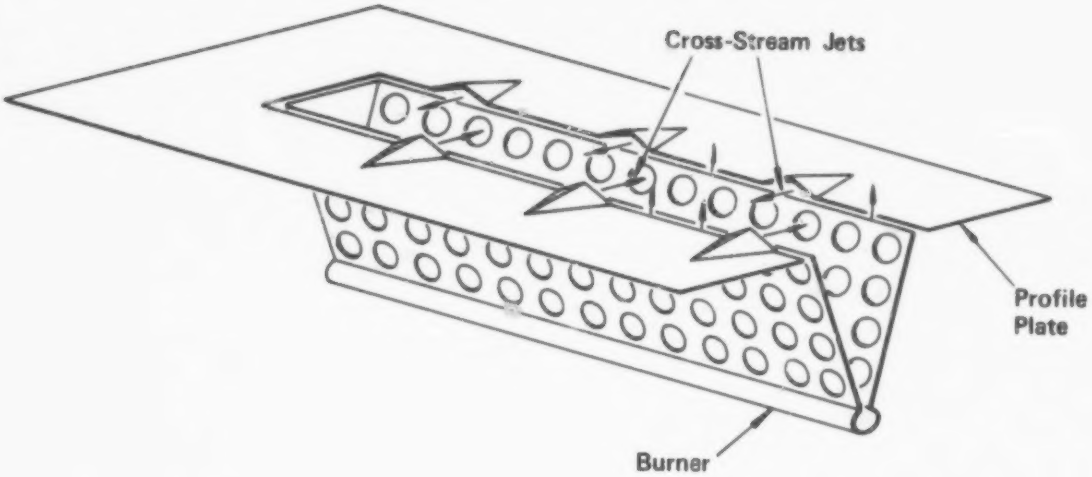

Cross-Stream Jets
Profile Plate
Burner

Figure 14-8. SUGGESTED APPLICATION OF CROSS STREAM JET MIXING TO PROFILE PLATES

232d

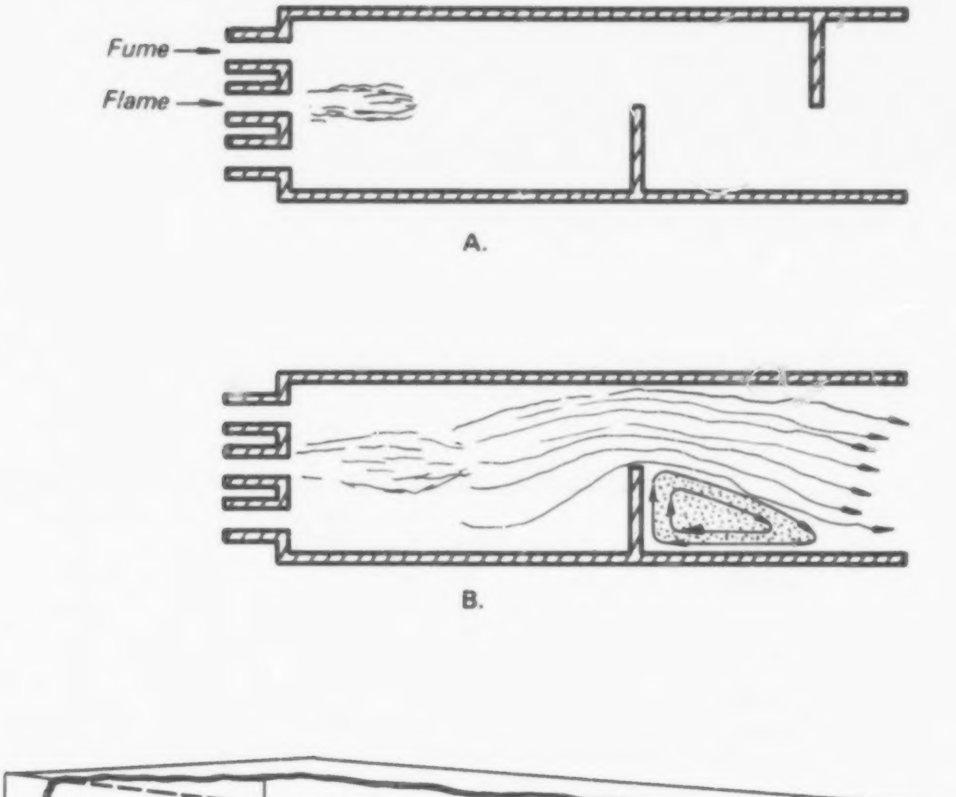

Fume
Flame
A.
B.
C.

Figure 14-9. BRIDGE WALL BAFFLES

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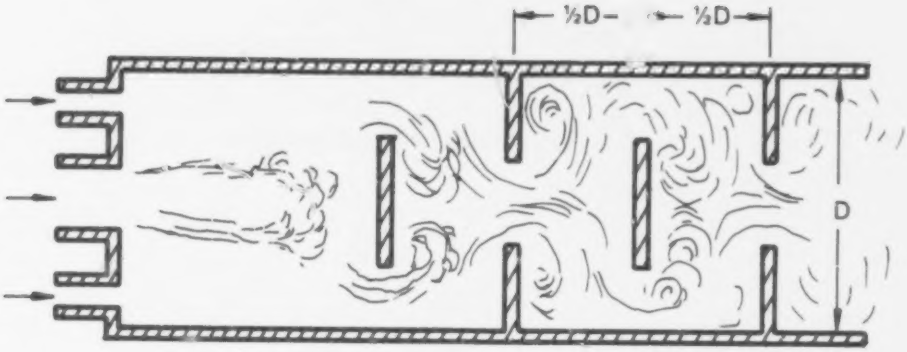
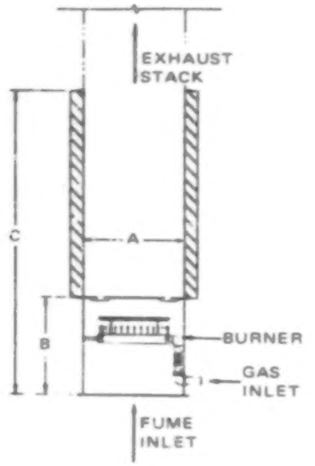
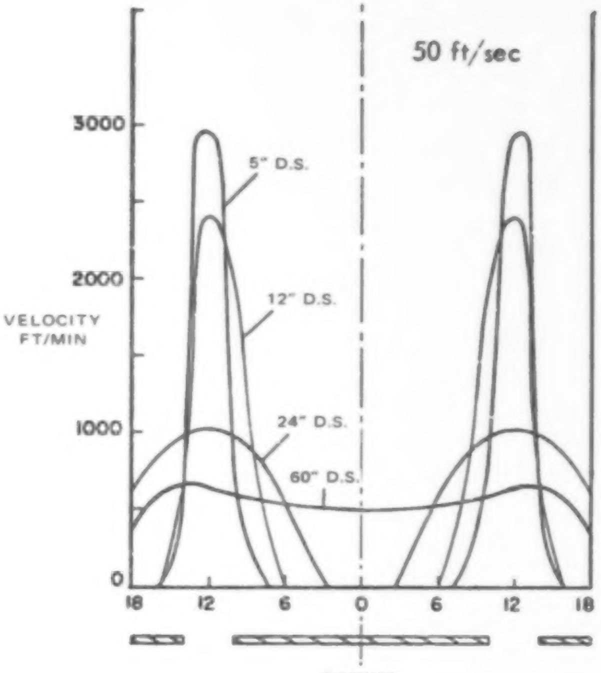


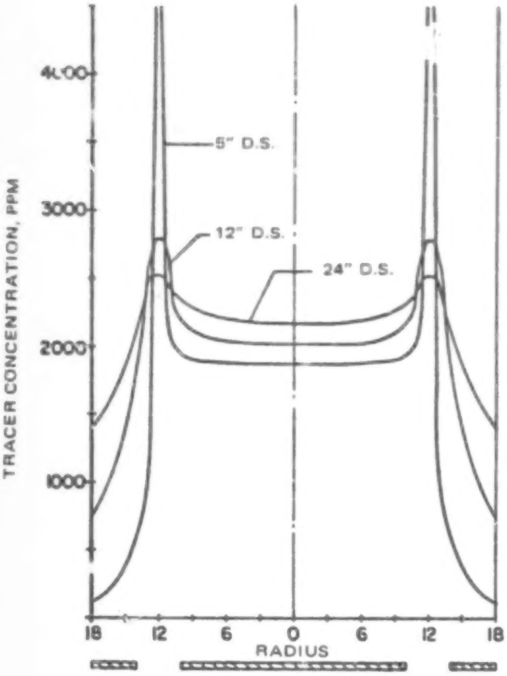
Figure 14-10. RING AND DISC BAFFLES


EXHAUST STACK
C
A
B
BURNER
GAS INLET
FUME INLET

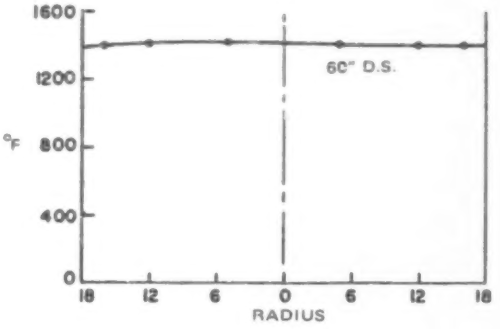
A. Burner Configuration



B. Velocity Development Downstream of Baffle



C. Tracer Concentrations Indicating Mixed Performance



D. Temperature Uniformity

Figure 14-11. PERFORMANCE OF AFTERBURNER WITH RING BURNER AND ANNULAR BAFFLE
(See Fig. 3-13)

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232g

COURTESY: *Surface Combustion Division – Midland Ross Corp. (Ref. 14-10)*

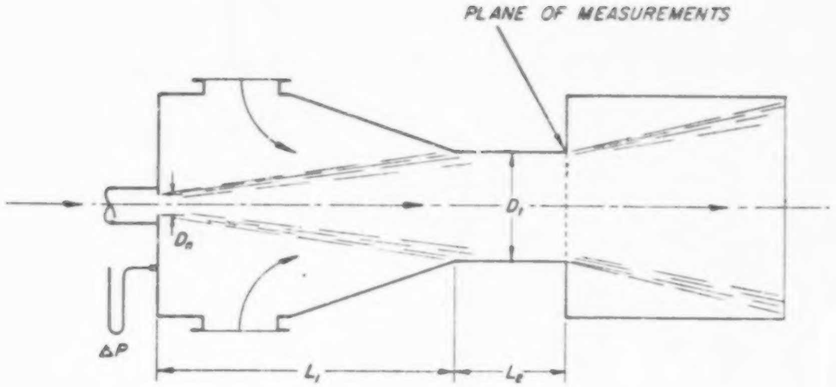

PLANE OF MEASUREMENTS
D_1
D_n
ΔP
L_1
L_2

Figure 14-12. JET INCINERATOR

COURTESY: *Surface Combustion Div. Midland Ross Corp.*

One observes that the undisturbed potential core flow has disappeared faster than would be predicted from the jet model; this is probably attributable to the higher level of turbulence generated by the flame beneath the opening. The jet model is based on the assumption of a smooth nozzle flow generating the jet. The baffle demonstrates good mixing as measured by temperature uniformity at the traverse taken 60 inches downstream from the opening, which would be about 26 L/ with respect to the jet, but is less than two diameters with respect to the chamber.

14.9 L-Turns and U-Turns

A right angle or reverse turn in the channel of the combustion chamber has mixing properties. There is considerable data on velocity distribution downstream from turns, but very little data on their effectiveness as mixing devices. Data of Reference 14-11 shows that from 6 to 10 diameters (or channel widths) distance downstream from the turn is necessary to re-establish a uniform velocity distribution; it can be assumed that this distance would also be needed to achieve a uniform condition of mixing. Thus, it would appear to be of limited value for the afterburner designer. More will be said about the use of L-and U-Turns in the next chapter on the design of the holding section.

14.10 Annular Throat: The Eductor Principle

Using the flame jet as the motive power, both mixing and pumping can be accomplished in an arrangement that makes the burner a jet eductor. Figure 14-12 shows an example of a burner supplied by one manufacturer, who claims that the arrangement is especially suited to handling dirty or liquid-droplet-containing streams, since the fume passages are not passed through the burner, and the use of an exhauster fan can be avoided. The burner uses outside air (which places an added heating burden on the system), and the blower used on this air supply provides the necessary motive power for the eductor. Figure 14-13 gives performance data on this incinerator (from Reference 14-5). References 14-1 and 14-15 give bases for designing the jet eductor. The jet must fill the throat, and the throat should have parallel walls for at least 2 diameters; for really effective mixing 8 or 10 diameters should be used. Care should be taken to be sure the flame is complete before it is used to entrain the fume; too rapid a quench of the flame will generate CO and aldehydes. Following the throat a diverging diffuser section may be used to recover the kinetic energy of the stream leaving the throat. Since such a diffuser must have an angle of divergence of no more than 15 degrees to maintain a completely filled duct, the length of the section is likely to make it uneconomical.

14.11 Side Entry of Flame or Fume

Mixing of flame and fume can be promoted by introducing the flame jet in the side of the afterburner, with the fume stream entering on the end and flowing along the axis. The desired pattern in the combustion chamber is a general random pattern of eddies. However, a purely radial or tangential entry is likely to develop one of the stable recirculating vortex patterns shown in Figure 13-12; neither of these patterns is desirable; the latter especially is undesirable, as it may set up a stable circulating flow that is likely to persist over the length of the whole afterburner, unless controlled by baffles. Model experiments with air in rectangular boxes (Reference 14-13) suggest that the burner or burners should be directed about 30° from the radial as they enter the chamber. (The model tests used 8 jets in two planes, with 4 directed for one rotation, the other 4 in counter-rotation; such an arrangement would be impractical for any but very large afterburners.) The model tests also showed that the axial entry of the fume stream should be at low velocity to allow good flame jet penetration.

Burners arranged for side entry may encounter flame impingement problems that could result in shortened refractory life and prematurely quenched flames. Flames designed to "fill the chamber cross section with a bath of flame" also run the risk of quenching the flame with the fume, thus leading to the formation of CO and aldehydes. It would be better on this score to complete the flame before it spreads across the chamber, thus using the flame products for mixing with the fume.

There is the alternate possibility of introducing the flame in the end of the chamber, and bring the fume in from the side. If this is done, the fume should be brought in radially and with a low velocity. If brought in tangentially at high velocity, the cool (and therefore denser) fume gases remain in a stable cyclonic flow adjacent the wall with little tendency to mix with the flame. Using the low velocity radial-entry fume arrangement the flame length can be kept short by using a burner having swirl in its air supply (see Chapter 13).

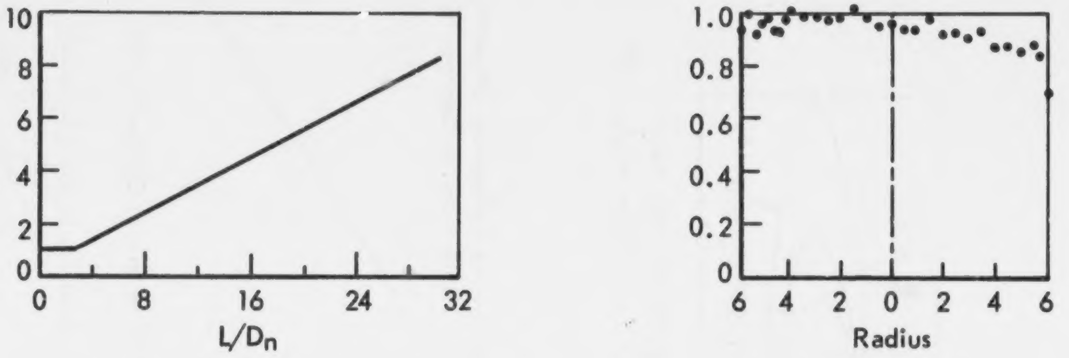
14.12 Mixing with Swirling Flames

Air model studies with the central jet (or burner) stream having swirl and the secondary air (or fume) stream having no swirl, mixing in open chambers, (Reference 14-16), in cylindrical (Reference 14-17) and in rectangular (Reference 14-18) chambers, show that the swirl significantly reduces the distance required for mixing as compared with a straight jet. In effect, the flame is made shorter and larger in diameter. Moreover the central potential core can be diminished or even eliminated. If the spreading jet impinges the wall to any appreciable degree, a central recirculating zone develops, as in Figure 13-7a. Results of the tests on a 1-inch diameter jet for no swirl and with a 45° swirl vane angle are shown on Figure 14-14. The effective length of the jet with the swirl has been cut in half.

14.13 Use of Cold Air and Water Models to Study Mixing Problems

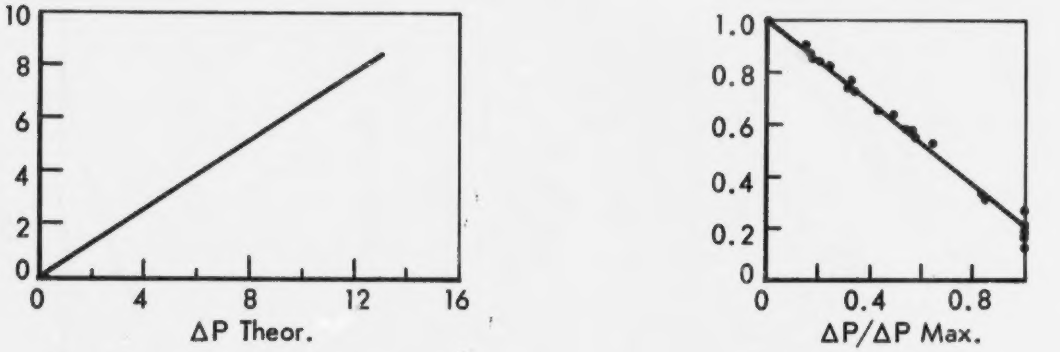
Where there is doubt that mixing conditions are adequate in a proposed afterburner design, the use of small cold air or water models is strongly recommended. Much information can be gained from relatively inexpensive equipment. Such a program is far less likely to be expensive than a campaign of cut-and-try on a full scale afterburner. The subject is covered more fully in Chapters 10 and 15.

Figures 14-13 and 14-14 follow



Maximum Entrainment Ratio vs Dimensionless Jet Length

Measured Mixing Efficiency



Maximum Observed Pressure vs Theoretical Pressure

Jet Incinerator Characteristic

Figure 14-13. PERFORMANCE OF AN EDUCTOR TYPE MIXER USED AS AN INCINERATOR

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COURTESY: *Surface Combustion Division – Midland Ross Corp* 14-12)

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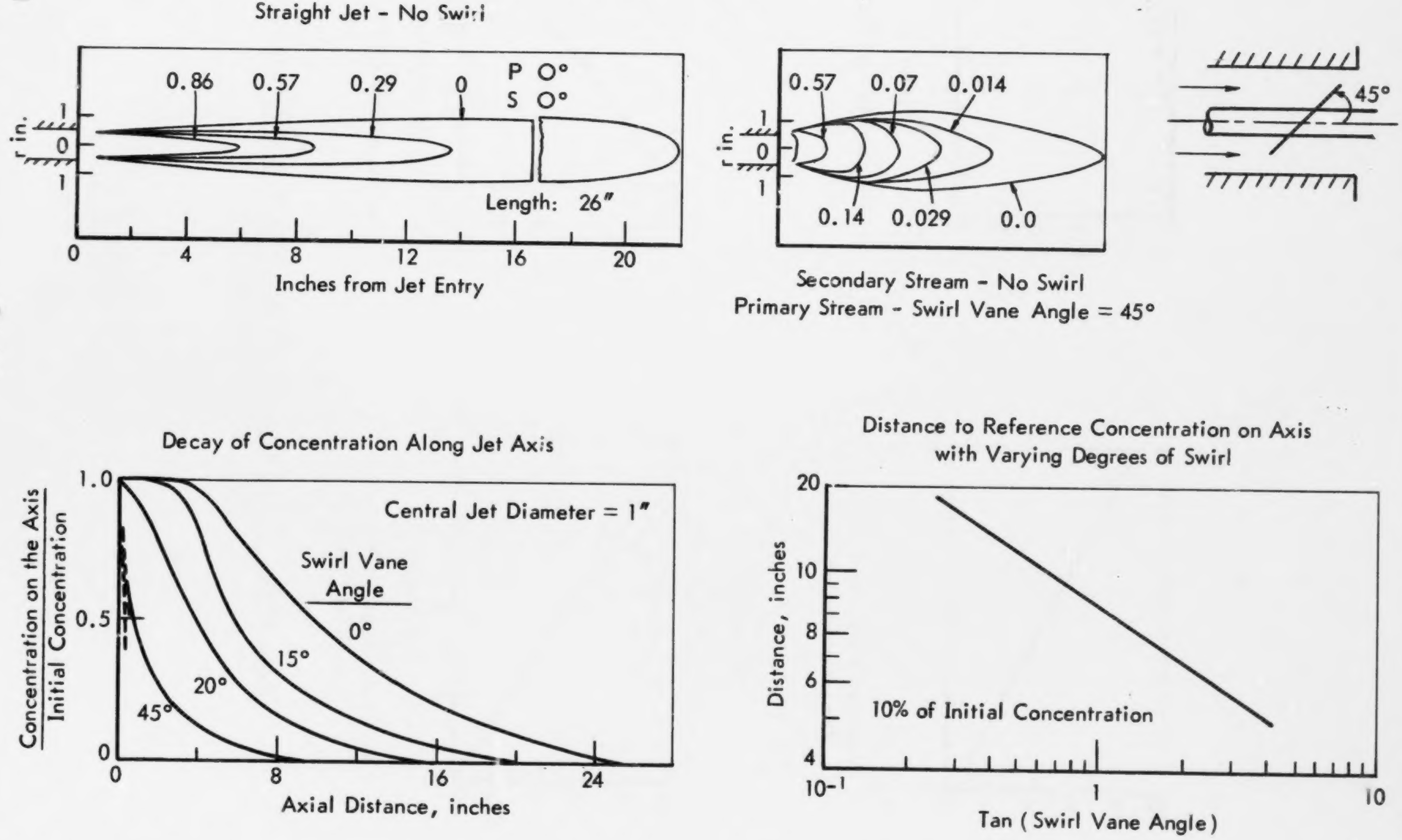

Straight Jet - No Swirl

Secondary Stream - No Swirl
Primary Stream - Swirl Vane Angle = 45°

Decay of Concentration Along Jet Axis

Distance to Reference Concentration on Axis with Varying Degrees of Swirl

Figure 14-14. CONCENTRATION PROFILES - JET MIXING WITH SWIRL

234c

Chapter 15. RESIDENCE CHAMBER

15.1 Introduction

Previously noted has been the division of internal processes of the afterburner into three parts: The burner, the mixing section, and the residence chamber. This is for convenience in analysis; in reality these processes overlap each other, and the burning of fuel and oxygen, the blending in of the dilute fume stream (only part of which can go through the flame without extinguishing it), and the holding at elevated temperature for a finite time to decompose the fume components cannot clearly be distinguished (Figure 12-3).

In this chapter it will be shown that a plug-flow condition is the ideal for promoting complete destruction of the noxious fume elements. Possibilities for achieving this ideal in practical designs will be reviewed. The effects of using common geometric shapes and chamber shapes will be noted, including the folded chamber, round vs. rectangular chambers, and the effects of baffles. Comments will be made on some exhaust stack arrangements, insofar as they affect the afterburner operation (i.e., not from the standpoint of dispersing effluents). Finally, there will be a discussion and recommendations on the use of cold air and water models as means of testing out and exploring different burner, mixing, and residence chamber alternatives prior to the construction of full scale units.

15.2 The Plug Flow Reactor

At some distance along the length of the afterburner the designer will conclude that mixing is essentially complete, that henceforward the space will be assigned to holding fume components and oxygen at elevated temperature long enough (~.1 - .3 seconds at 1400° - 1500°F for most pollutants) to ensure that the reaction will be completed within the required limits. The last chapter took considerable pains to show that mixing of flame and fume is in reality very difficult, and that a truly uniform distribution of materials and that even profiles of temperature and velocity are unlikely. Let us assume that there is a known distribution of temperature and flow velocity at the entrance to the residence chamber, as determined by experimental measurements with thermocouples and pitot tube, or as estimated with the help of Chapters 13 and 14. Let us explore the consequences of this distribution.

Figure 15-1 represents a hypothetical traverse of velocity and temperature across an afterburner at the start of a long, straight residence chamber, of length A to B. Having entered this section in a nearly-well-mixed state, there is very little velocity gradient in the central part of the chamber to generate the eddies that would promote further mixing. There may be some residual turbulence from the previous eddy motion; otherwise, any eddy mixing or further turbulence must come from wall drag, and the resulting velocity gradients next to the wall. That is, the flow will approximate turbulent flow in a pipe. In general, mixing under conditions of turbulent flow in a straight pipe are not very good; commonly 40 diameters is allowed to even out any initial mal-distribution. (This depends somewhat on Reynolds Number, of course.) Since the usual afterburner residence chamber has a length which is typically from 3 to 5 times the diameter or width, very little mixing can be expected while traversing the length of the chamber. As seen in Figure 15-1, gas entering section A at temperature T and velocity V will pass along a stream tube and leave at B with very little change. The situation shown in the sketch is a favorable one - the higher velocity element of gas also has the higher temperature, which promotes a faster conversion to offset the shorter residence time. A much less favorable situation would have a high velocity portion of the stream passing through at low temperature; this will happen if a slipstream of fume should bypass around the burner without subsequently mixing with material passed through the flame. Clearly, the most favorable condition exists when both velocity and temperature profiles are flat. This condition gives what is called a "plug flow" reactor; that is, the gas moves along as though it were a solid plug.

15.3 Plug Flow Reactor Residence Time

Many pollution control agencies require that an afterburner have a design that will hold the fume stream at a specified temperature for a specified "residence time," with temperature and time values fixed according to the composition of the fume stream or the type of service. Strictly speaking, this time should apply to the plug flow reactor section. Conservatively, the time and space used by the burner and the mixing of burner combustion products with the fume should precede and be independent of this residence time. In estimating the residence time in the plug-flow reactor section, the average velocity of the mixed gases should be computed, based on the chamber pressure and temperature (see equations given in Chapter 13 for calculating the gas density), and finding:

$$t = \frac{\text{Length of Plug Flow Reactor}}{\text{Average Velocity}}$$

For the computer-minded, an approximate estimate of the conversion of the pollutant components should be possible, using the plug flow reactor concept as applied to each flow element across the channel cross section. Such a calculation can be only approximate, since

a) Despite the arguments given above, there is indeed some cross-stream mixing, hence all the gas coming out the opposite end of the stream tube will not have the same temperature or will have experienced the same residence time.

b) Reaction must be describable as first-order for the pollutant, which also requires that there be an excess of oxygen at all times.

By estimation or by physical measurement one can establish a profile of initial concentration C_o of each of the pollutant elements, such as CO, methane, hexane, etc., over the section A-A. The area may be subdivided into equal elements, dY. Chapter 12 has given reaction-rate data for some pollutant components, and many other materials can often be treated as though they were one or another of these components.

Figure 15-1 follows

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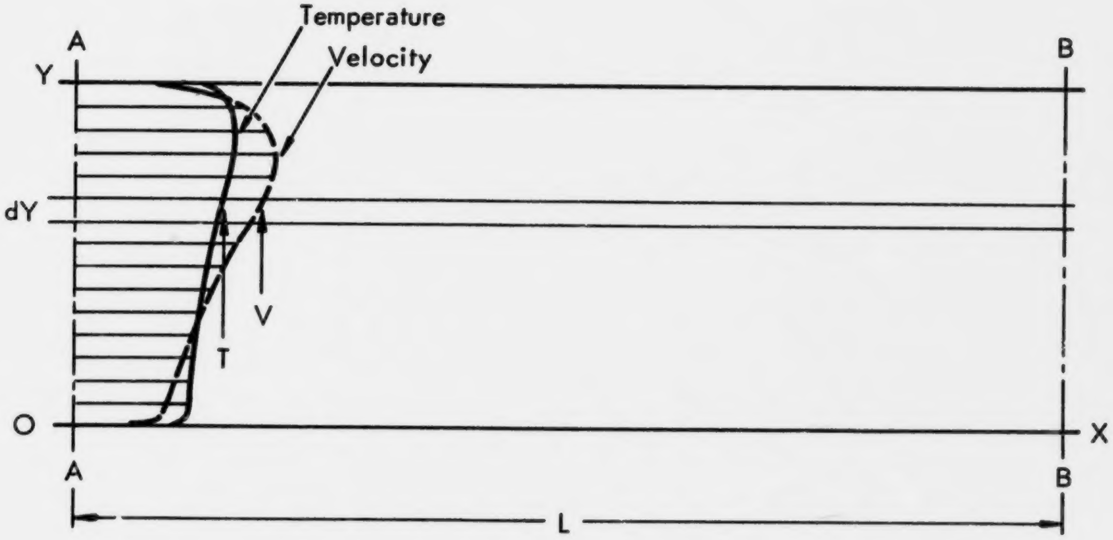

Temperature
Velocity
A
B
Y
dY
O
X
A
B
T
V
L

Figure 15-1. THE PLUG FLOW REACTOR

The time-of-flight of the gas entering the chamber at A-A and leaving at B-B can be found for each area element from the velocity distribution,

$$t = \frac{L}{V}$$

For each element, the residual concentration at B-B can be found,

$$C = C_o \, e^{-kt}$$

where

$$k = A \, e^{-E/RT}$$

E is the activation energy, R the gas constant, T the absolute temperature. Values of A and E are given in Chapter 12. The total conversion, and the remaining residuals may be found by summing over the area elements.

15.4 Chamber Shapes

The simplest shape, and functionally the best for most closely achieving the plug flow reactor condition, is the straight cylinder. A rectangular cross section with a straight axis would be a second, but generally acceptable choice - there is some tendency for the corners to have a lower flow velocity. The rectangular shape is easier to construct, especially for large sizes of afterburners.

Some afterburner manufacturers have endeavoured to save space needed for the afterburner by bending the residence chamber back on itself, or by using an ell-shaped space. This is not good practice. On the inner face of the wall, downstream of the bend, and in the corners, flow separations occur. The stream does not follow the wall, and large, recirculating eddies are formed. These eddies remain stably in place, become stagnant pockets of burned-out gas, and do not contribute to the pollutant oxidation reaction. Figure 15-2 shows the volume occupied by these recirculating eddies. Their volume should be subtracted from the effective volume of the residence chamber and may reduce total residence time by as much as 30%. The bends are also unsatisfactory in that they tend to give a highly non-uniform velocity distribution, even outside the recirculating pockets, so that some of the material is bypassed with less than its proper residence time in the chamber, some is held for more. The arrangement is far from that of a plug-flow reactor (Reference 15-1). Figure 15-3 shows some estimates of the effect of an L-bend on the residence time and velocity distribution of material traveling along a streamline with a velocity differing from the average.

Other manufacturers have used the bends, or baffles as a means of improving the mixing or contacting in the residence chamber. Such mixing or contacting should be carried out before the residence section.

Chapter 14 mentioned the checker wall or screen in connection with devices for improving mixing. Its use at the entrance to the residence section can be beneficial, especially if it follows a U-bend or L-bend, or if there is a poor velocity distribution at the entrance. The checker generates a pressure drop that redistributes the flow and gives a more even velocity distribution downstream. It does generate additional turbulence, but this is a relatively small scale turbulence which dies away a short distance downstream (Reference 15-2).

15.5 Comments on Stack Arrangements

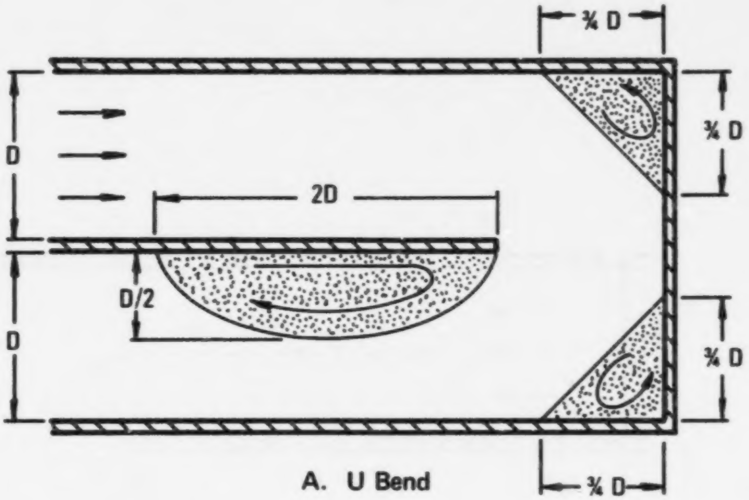
The residence section should terminate in a channel of reduced cross section, especially if the chamber axis is vertical. The flow velocity in the residence section is usually low, and a large vertical stack under low flow conditions may develop recirculating flow patterns, aided by the bouyancy of the hot gases and the cooling on the walls. Under no circumstances should the interior of the chamber "see" the open sky, since the sky is a very effective "sink" for radiative cooling of the chamber walls. Walls of the chamber should be insulated, as otherwise material passing along the cooled surfaces will escape without being oxidized.

15.6 The Use of Cold Air and Water Models

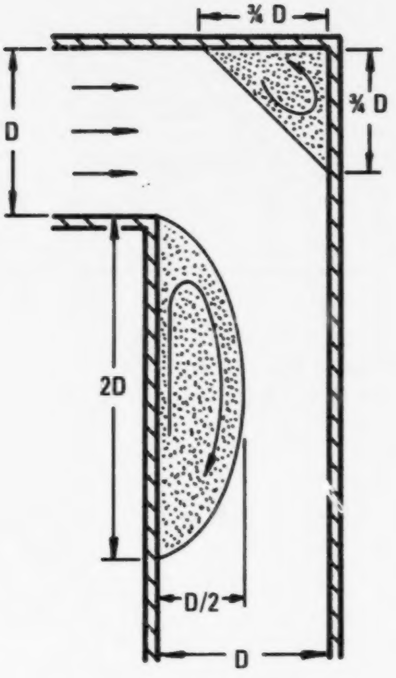
As mentioned in Chapter 14, and discussed in Chapter 10, where there is doubt about the effectiveness of the mixing between flame and fume streams, and also where there is uncertainty over the success in providing conditions at the entrance to the residence section which will approximate the desired plug-flow conditions, cold air or water models should be considered. Such a model can be set up in the laboratory for a small fraction of the cost of constructing a complete afterburner. It can be built of sheet metal or wood, although "Plexiglass" is much to be preferred, since it allows visual probing and observation with smoke tracers and dyes. Mixing effects can be studied while blending warm air or water into room-temperature air or water. Air is to be preferred here, as the density is much closer to that of the hot gas in the afterburner, and the Reynolds Number will likewise be more representative. However, water models can be especially useful where blending or mixing with colored dyes is desired. Sensitive laboratory-type thermocouples and pitot tubes must be used with the air models to obtain satisfactory traverse readings of temperature and velocity distribution. Sections 10.6 and 10.7 should be consulted for more details.

In preserving the scale validity between air model and full scale afterburner, corrections should be made to account for the differences in density of the model and the prototype. Thring (Reference 15-3) gives a method for this correction, based on the relative momentum of the stream being considered.

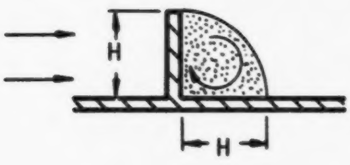
Figures 15-2 and 15-3 follow


¾ D
¾ D
D
2D
D/2
D
¾ D
¾ D

A. U Bend

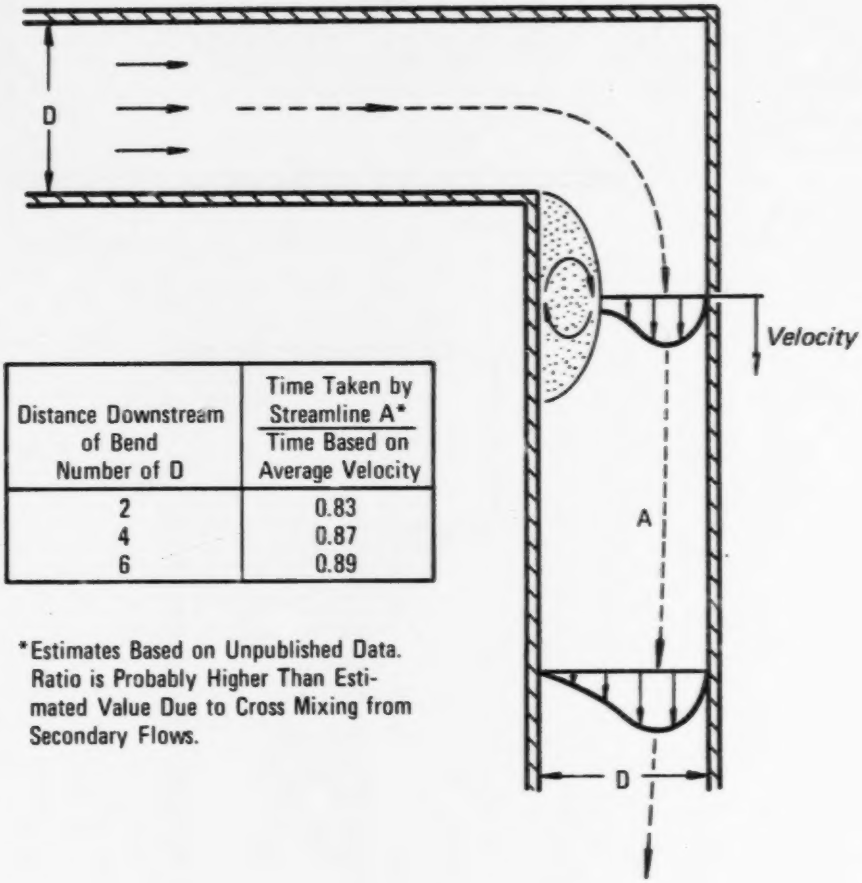

¾ D
¾ D
D
2D
D/2
D

B. L Bend



C. Wall Baffle

Figure 15-2. BLOCKAGE EFFECTS OF TURNS AND BAFFLES ON RESIDENCE CHAMBER



Distance Downstream of Bend Number of D	Time Taken by Streamline A* / Time Based on Average Velocity
2	0.83
4	0.87
6	0.89

*Estimates Based on Unpublished Data. Ratio is Probably Higher Than Estimated Value Due to Cross Mixing from Secondary Flows.

Figure 15-3. EFFECT OF L-TURN ON VELOCITY DISTRIBUTION AND RESIDENCE TIME DISTRIBUTION IN CHAMBER

Chapter 16. MECHANISMS AND RATES OF REACTIONS IN AFTERBURNER CATALYSTS

16.1 Mechanisms

In an afterburner catalyst, reactants flow through the bed (or matrix) and reaction products are carried away by the flow of a waste (gas) stream. (See descriptions of the structures of such catalysts in Section 4.3.1) The reactions which convert reactants to product molecules, however, take place at discrete catalytic sites distributed throughout a thin, porous substrate layer coated on the surface of the matrix. The overall process of converting reactants to products in the flowing stream involves a number of sequential steps including transport processes between gas stream and catalytic sites as well as the actual steps in the reaction at the catalyst sites. These steps can be listed:

1) Mass transfer of combustibles and oxygen to the external surface of the catalyst,

2) Diffusion of combustibles and oxygen into the pores of the substrate layer,

3) Adsorption of combustibles and/or oxygen on the active catalytic sites,

4) Reaction (oxidation) at the active site,

5) Desorption of combustion product molecules from the catalyst sites,

6) Diffusion of combustion products through the porous substrate to the external surface,

7) Mass transfer of combustion products from the catalyst surface into the flowing waste stream.

The oxidation reactions taking place in Step (4) are highly exothermic, liberating considerable heat at the active sites. This in turn must be transfered to the waste stream by thermal conduction through the porous layer and by convective or conductive transport from the catalyst surface to the flowing gas stream.

The qualitative effects of these sequential processes are that concentrations of combustibles are lower within the catalyst, at the active sites, than in the gas stream flowing over its surface (and that combustion product concentrations and the temperature are higher within the catalyst). The concentration and temperature differences which develop, provide driving forces for the various diffusion and heat and mass transfer steps in the overall sequence. Their quantitative importance depends on the relative magnitudes of rate constants describing each of the steps in the sequence.

A complete description of the rate of the overall process would require a knowledge and description of each of the seven steps in the sequence. The present state of knowledge (and available data) on afterburner catalysts does not permit such a detailed breakdown. However, sufficient data are available to allow independent description of the rates of convective mass transfer processes (Steps 1 and 7) taking place external to the catalyst, and of the net process taking place within the porous catalyst structure (Steps 2 through 6). The theory(6-5,7) and some basis also exists for treatment of the interaction of diffusion and reaction within the porous catalyst substrate. This interaction has not been treated explicitly in the data correlations of this report. However, sufficient calculations using the theory have been made to make it evident that under normal design rates for these catalysts, the effective reaction rate is impeded by slow diffusion of reactants through the porous substrate--to such an extent that only that fraction of the catalyst near the surface of the thin porous substrate layer (~0.005 inch in thickness) is being used effectively to catalyze the oxidation reactions. Under these conditions the "chemical rate" (i.e., the net rate of Steps 2 through 6) is proportional to the external surface area of the catalyst matrix rather than to the volume of the porous catalyst. A second consequence of the inhibition by diffusion is the reduction in the temperature dependence of this rate to approximately half what it would be without diffusional interaction. However, the entire region of practical rates appears to lie within the region in which this lower (constant) temperature dependence prevails. Thus, a pragmatic description of the "chemical rate" is possible without the need for separate calculations of "effectiveness factors" to account for the interaction of intracatalyst diffusion and chemical reaction.

Heat transfer between catalyst and gas stream is important because it determines the internal temperature of the catalyst, the temperature at which chemical reactions take place at the active sites. To be most useful, any correlation of rates of chemical reactions should be in terms of this internal temperature rather than the temperature of the gas stream flowing over the catalyst surface. Calculations for typical design conditions indicate negligible difference in temperature between catalyst surface and internal points within the catalyst (the solid thermal conductivity is high and the layer of porous substrate is thin). However, near the inlet end of the catalyst bed, when the hydrocarbon content of the waste stream is high, the temperature difference between gas stream and catalyst surface can be significant (25 - 100°F).

The same steps are involved in the use of heterogeneous catalysis in the chemical or petrochemical industries, where the objective is to manufacture a valuable product; as in catalytic afterburning where the objective is to destroy an unwanted reactant. However, the relative importance of the steps may be different since conditions and objectives are different. In industrial heterogeneous catalysis, typically 10 - 100 cu ft of catalyst bed are used for each 1000 SCFM gas stream to be reacted, whereas 1 - 2 cu ft/1000 SCFM is typical in afterburners. If the comparison is made on the basis of actual porous catalyst in the bed; the rates per unit volume of active catalyst are 2 - 3 orders of magnitude higher in afterburners than in conventional, industrial applications of porous catalysts. To a large extent the higher rates are possible in afterburners because the desired products are drastic (total) oxidation products and selectivity to these products is high and is not decreased by conditions leading to high rates. In industrial catalysis, operation is normally limited to a relatively narrow region of moderate reaction rates in which the catalytic reaction to form the desired product (of partial oxidation, partial hydrogenation, etc.)

is strongly favored over other possible reactions of the same reactants. As a consequence of the extremely high reaction rates of afterburner catalysts, afterburners represent one of the few applications of porous heterogenous catalysts in which the rate of heat and mass transfer from the flowing reactant stream to the catalyst surface have a strong effect in determining the overall reaction rate.

16.2 Rates of Reactions

16.2.1 Empirical Observations

The oxidation of most hydrocarbon (and oxygenated hydrocarbon) species over supported noble metal catalysts under typical afterburner design conditions is complete when sufficient oxygen is present. The reaction stoichiometry may be written:

$$C_nH_m + (n+\frac{m}{4})O_2 \longrightarrow \frac{m}{2}\,H_2O + n\,CO_2 \qquad (16\text{-}1)$$

with CO_2 and H_2O as the products of the catalytic reaction.

The rate of oxidation (reaction) of a given hydrocarbon species over the catalyst is proportional to its concentration in the gas stream and independent of the concentration of oxygen in the gas stream (provided there is at least 2%v oxygen in excess of that required for complete oxidation of gaseous combustibles). For oxidation of waste streams containing several oxidizable species it is usually assumed that each species is oxidized independently with its own rate.

The rates of oxidation (under the same conditions) for similar catalysts prepared on supports of different geometric surface areas are proportional to the geometric surface area of the matrix (rather than the volume of catalyst).

16.2.2 Rate Expressions

The above empirical observations are consistent with the expression for the local rate of reaction at a point within an afterburner catalyst bed:

$$\text{Rate}\left(\frac{\text{moles}}{\text{sec-ft}^3}\right) = -\,V_g\,\frac{dC_{HC}}{dZ} = \bar{k}_s \cdot \frac{s}{v} \cdot C_{HC} \qquad (16\text{-}2)$$

where C_{HC} is the concentration of a particular hydrocarbon species (lb-mole/ft^3)

V_g the superficial velocity of gas through the bed (ft/sec)

Z the length dimension in the direction of flow (ft)

$\frac{s}{v}$ the ratio of geometrical (external) surface to volume of the catalyst matrix (ft^2/ft^3)

$\bar{k}_s$ is the effective overall rate constant for oxidation (ft/sec).

The effects of temperature, flow conditions through the catalyst and catalyst geometry are consistent with $\overline{k}_s$ representing two rate-limiting processes in series:

1) Mass transfer of reactant hydrocarbon from the waste stream to the catalyst surface,

2) An effective chemical reaction within the catalyst, localized near its external surface.

The rate of each process is first-order in reactant hydrocarbon concentration. The overall rate constant $\overline{k}_s$ can then be represented in terms of the rates of these two processes:

$$\frac{1}{\overline{k}_s} = \frac{1}{k_{mt}} + \frac{1}{k_{chem}} \qquad (16\text{-}3)$$

16.2.2.1 Chemical Rate Constants

The chemical rate constant, k_{chem}, which represents the rate of the net processes taking place within the porous **substrate layer**, is determined by the chemical composition and microstructure of the active metal crystallites distributed in that layer and by the microstructure of the porous substrate. These factors are in turn determined by the method used in preparing the catalyst. The chemical rate constant depends also upon the species being oxidized, although for some catalysts, hydrocarbon species covering a relatively wide range of molecular structures have quite similar rate constants (see Section 4.2.2). It has a relatively strong dependence upon temperature, which can be represented over a reasonable temperature range by the Arrhenius form:

$$k_{chem} = A \exp\left(- E_a/R_g T_s\right) \qquad (16\text{-}4)$$

where A is the pre-exponential factor depending upon catalyst type and species being oxidized

E_a is the activation energy

R_g the gas constant

T_s the absolute temperature of the catalyst.

At lower temperatures the behavior of the overall rate constant, $\overline{k}_s$, is limited by k_{chem} and its value approaches that of k_{chem}. It is from data obtained at lower temperatures, where the overall rate shows strong temperature dependence, that one is best able to derive correlations for k_{chem}. Data concerning the chemical activity or rate constants of afterburner catalysts are usually considered proprietary by the equipment manufacturer. Consequently, few such data are generally available for development of correlations for k_{chem}. Some data (see

Figure 4-5) have been published on relative overall rate constants for all-metal catalysts. These data provide some basis for estimation of an average value of E_a for this catalyst and for estimating relative values (ratios) of A factors for the hydrocarbons included in the figure. However, absolute values of A factors cannot be derived from the data presented. Data, such as that in Table 4-2 on ignition and operating temperatures give some indication of values of k_{chem}. A few additional data have also been published in the form of plots such as that reproduced in Figure 4-2, giving conversion versus temperature for a specific hydrocarbon/catalyst combination under defined operating conditions. These data are the most useful in developing correlations--but are very limited in number.

16.2.2.2 Mass Transfer Coefficients

The other rate limiting factor, k_{mt}, is determined by the geometry (matrix structure) of the catalyst bed, and by the hydrodynamics of flow of the waste stream through the catalyst bed. It has a relatively slight dependence upon temperature and upon the nature (molecular size and shape) of the hydrocarbon being oxidized. The overall rate constant, $\bar{k}_s$, approaches the value of k_{mt} as a limit at high temperature when the value of k_{chem} has become large relative to k_{mt} (see Figure 4-5). Data on performance of afterburner catalysts at high temperatures provides one source of information for developing correlations of k_{mt}. However, a much broader base of information is available from the heat transfer literature, if the analogy between heat transfer and mass transfer under the same hydrodynamic conditions is used. Theoretical calculations or experimental studies have been made to develop heat transfer correlations for many matrix structures either identical with or similar in geometry to those used as supports for afterburner catalysts[16-2)3)4)6)]. These correlations provide directly a basis for obtaining heat transfer coefficients to allow calculation of the temperature difference between gas stream and catalyst surface. Indirectly, they allow independent development of correlations for gas-to-solid surface mass transfer coefficients by analogy.

The results of heat transfer studies on a particular matrix geometry can usually be expressed in a correlation of the form:

$$N_{Nu} = \frac{hL}{k} = f\,(N_{Re},\ N_{Pr}) \qquad (16\text{-}5)$$

where N_{Nu} is the dimensionless Nusselt number, hL/k

h is the solid-to-gas heat transfer coefficient (Btu/hr-ft^2-°F)

L is a characteristic dimension of the matrix, usually "hole diameter" or effective diameter of pellet or wire (ft).

k is the thermal conductivity of the gas (Btu/hr-ft-°F).

N_{Re} is the dimensionless Reynolds' number, $LG/\varepsilon\mu$

G is the superficial mass velocity of the gas through the matrix (lb/hr-ft^2)

μ is the viscosity of the gas (lb/ft-hr)

ε the fraction void of the matrix.

N_{Pr} is the dimensionless Prandtl number, $C_p\ \mu/k$

C_p is the heat capacity of the gas (Btu/lb-°F)

f is the correlating function, the form of which depends on the matrix geometry.

The analogy between heat and mass transfer, as used here, utilizes such a heat transfer correlation to develop the mass transfer correlation, which is written:

$$N_{Sh} = \frac{k_{mt}\ L}{D} = f\ (N_{Re},\ N_{Sc}) \tag{16-6}$$

in which for a given matrix, f is the same function of Reynolds' number as in Equation (16-5), but with the Schmidt number, N_{Sc}, replacing the Prandtl number where it appeared in the function in Equation (16-5).

N_{Sh} is the dimensionless Sherwood number, $k_{mt}\ L/D$

k_{mt} is the gas to solid mass transfer coefficient (ft/hr)

D is the gas diffusivity (ft^2/hr)

N_{Sc} is the dimensionless Schmidt number, $\mu/\rho D$

ρ is the gas density (lb/ft^3).

This analogy has been used to develop correlations for k_{mt} for all of the matrix structures in use as afterburners. These are presented in Chapter 17. Where possible these have been compared with mass transfer data obtained from high temperature performance of afterburner catalysts.

The matrices used as catalyst supports fall into two general classes with respect to their hydrodynamic interaction with the fluid flow. The honeycomb types are essentially bundles of parallel tubes with developing (or in the limit fully developed) laminar flow through these tubes. The all-metal and Oxycat configurations approach more nearly the case of cross flow through a tube bundle, or flow through a screen or mesh array.

Kays and London[16-2] present analytical solutions for the Nusselt (or Sherwood) numbers for fully developed laminar flow through tubes of various cross-sectional geometries and the results of numerical solutions of the exact governing equations for development of laminar flow in smooth tubes of circular cross section. The numerical solution was curve-fit and combined with the analytical solutions for fully developed flow to give a general semi-theoretical correlation for mass (or heat) transfer coefficients in honeycomb matrices. This is:

$$N_{Sh} = B\ [1 + 0.078\ L/X\ N_{Re}\ N_{Sc}]^{0.45} \tag{16-7}$$

where L the length dimension used in N_{Sh} and N_{Re} and also appearing explicitly is four times the hydraulic radius of the individual tube in the matrix.

X is the length of the uninterrupted tube (the thickness of a layer of catalyst matrix material in the catalyst element).

B is a constant depending upon the shape of the tube cross section-- B = 3.66 for circular tubes, 2.35 for triangular tubes.

The mass transfer coefficient used in N_{Sh} is the length-mean coefficient over the length, X. The form of the equation, with respect to dependence of k_{mt} on L, X, and G was supported by data obtained by D. M. Sowards of the Industrial and Biochemical Department at E. I. duPont. On the basis of comparison of the equation with those data, the equation was modified to contain a somewhat higher dependence on the tube length between interruptions in the developing boundary layer.

$$N_{Sh} = B\,[1 + 0.095\ L/X\ N_{Re}\ N_{Sc}]^{0.45} \qquad (16\text{-}8)$$

(This empirical adjustment is not without theoretical basis, since the original equation was developed for flow through tubes with smooth walls. Surface roughness in the catalyst matrix material may well delay boundary layer development). It should be noted that as tube length is increased or diameter decreased the Sherwood number approaches its asymptotic value, B, for fully developed laminar flow. With decreasing length to diameter ratio of the tube (X/L) and increasing Reynold's number the second term within the brackets can make a significant contribution to the value of the Sherwood number. In a typical design with 1/8-inch honeycomb used in 1-inch thick layers, the Sherwood number may be 13 compared with 3.66 for fully developed flow in the tubes. Thus, for this catalyst, frequent interruption of the boundary layer and high mass velocity contribute to higher values of k_{mt} and improved conversion over the catalyst in the region of mass transfer control. On the other hand, under typical design conditions for catalyst supported on 8 C/inch Thermacomb, the Sherwood number may be only 2.8 compared with the value of 2.35 for fully developed laminar flow. With the smaller tube size of this latter catalyst, the mass transfer coefficient becomes almost independent of gas flow rate through the catalyst.

For the Oxycat and metal-ribbon supports, which approximate cross-flow configurations, correlations of the Grimison form[16-4] are recommended for Sherwood and Nusselt numbers.

$$N_{Sh} = a\ N_{Re}^{\ b}\ N_{Sc}^{\ 0.333} \qquad (16\text{-}9)$$

where a and b are empirical constants, with values between 0.5 and 0.6 for banks of staggered tubes. The length dimension used in Reynolds', Sherwood and Nusselt numbers is an equivalent diameter, D_e, equal to the perimeter (of the tube or ribbon) divided by π. The void fraction used in the Reynolds' number is that for the minimum flow cross section (rather than the average ε for the entire matrix).

Correlations of the type given in Equation 16-9 have a history of successful use in correlating and predicting heat transfer coefficients for banks of circular tubes. The extension here to arrays of elements of other than circular cross section is an extrapolation, but using accepted methods. Most of the data on which the Grimison correlations are based were obtained at relatively high Reynolds' numbers ($>5 \times 10^3$). In selecting the coefficients for Equation 16-9, data for heat and mass transfer coefficients to arrays of wire screens at low Reynolds' numbers were also taken into account16-2)6). The correlation given in Chapter 17 for the metal ribbon catalyst gives reasonable agreement with these data. Unfortunately, no direct confirmatory data from catalyst performance are available for the correlations proposed for either the Oxycat or the metal ribbon matrix. (Although presumably the manufacturers of these catalysts do have such data, they consider them to be proprietary.)

A third general type of form for heat and mass transfer correlations for catalysts is that for conventional beds of catalyst particles (generally spherical or short cylindrical pellets packed randomly). Beek16-1) reviewed available data and recommended the form:

$$N_{Sh} = 2.42\, N_{Re}{}^{1/3}\, N_{Sc}{}^{1/3} + 0.129\, N_{Re}{}^{0.8}\, N_{Sc}{}^{0.4} + 1.4\, N_{Re}{}^{0.2} \qquad (16\text{-}10)$$

This equation, like Equation 16-9, is a strictly empirical correlation, but does take into account with the three separate terms, the shifting dependence of the transfer coefficient with Reynolds' number as Reynolds' number increases. In Equation 16-10, the length dimension used is the effective particle diameter, and the Reynolds' number is defined as:

$$N_{Re} = \frac{d_p\, G}{\mu}$$

using only the superficial mass velocity, G, without including the fraction voids in the bed as a factor.

Chapter 17. DESIGN METHODS FOR CATALYTIC AFTERBURNERS

17.1 Introduction

The objective of this chapter is to outline and present a method for calculating the oxidation performance of a given volume and geometry (length and cross-sectional area) of catalyst bed or matrix. These calculations of performance are based on a minimum of experimental data obtained from measurements of catalyst oxidation performance (in correlated form) augmented by theoretical and semi-theoretical predictions of heat and mass transfer coefficients and friction factors. The model used in formulating the method is sufficiently realistic in representing the processes actually taking place within the catalyst bed that effects of gas velocity, temperature, concentration of hydrocarbon in the waste stream, and catalyst bed length are well represented in their combined effects on performance.

The calculation of catalyst performance is not the only problem in design of a catalytic afterburner. The mechanical and geometrical design of the preheat-mixing section to provide a uniform flow of preheated waste stream through the catalyst bed is also important. The considerations and design methods involved for the preheat-mixing section are similar to those discussed in Chapter 14 for thermal afterburners. Although, in general, the ratio of combustion products to fume stream to be mixed in the preheat section is lower for catalytic than for thermal afterburners, it lies within the range discussed in that chapter.

The performance calculation described here is aimed explicitly at the performance of fresh catalyst (where activity has not been reduced by poisoning or surface deposits). Its extension to calculations for aged, poisoned, or suppressed catalyst would be straightforward, but would require significantly more information than is currently available concerning both the nature and quantitative extent of catalyst fouling. It was pointed out in Section 4.2.4 that over-deisng of the catalyst bed to allow for decline of catylst activity is an important aspect of overall afterburner design.

To a large extent, in current practive the catalyst volume and geometry are specified on the basis of subjective evaluation and extrapolation from the results of pilot experiments. At least one catalyst manufacturer has developed a computer program for performance calculations similar to that described here. His method does not separate mass transfer from chemical rate effects and hence does not describe the influence of gas velocity on oxidation performance.

All of the equipment and catalyst manufacturers have available correlations (at least graphical) of the effects on pressure drop of temperature, gas velocity and bed length for the catalyst(s) they use. The method described here incorporates such pressure drop calculations into the overall performance calculation.

17.2 Performance Calculation Method

The method requires that the composition of the waste stream entering the catalyst bed be characterized in terms of the concentrations (lb moles/lb waste gas stream) of n combustible species (or groupings of species). The heat of combustion (Btu/lb mole), "chemical" oxidation rate constant as a function of temperature as defined in Chapter 16, **and diffusivity** must be supplied for each of the n species. Performance is calculated as the fraction of each of the n species remaining unoxidized after passing through a specified bed length with specified inlet temperature (preheat) and mass velocity.

Differential equations describing chemical and physical processes in the catalyst bed are formulated using the models of these processes described in Chapter 16. These equations are formulated using normal methods used in developing reactor design equations for industrial catalytic reactors. The only major assumptions made are that the catalyst bed operates adiabatically and at steady-state, and that flow rate of the preheated waste stream is uniform over the bed cross section.

The governing differential equations for the axial changes of concentrations, temperature and pressure of the flowing gas stream can be written (see Nomenclature at end of this chapter):

$$- G\frac{dC_i}{dZ} = \bar{k}_{s,i}\,\frac{s}{v}\,C_i\,\rho_g \qquad (i = 1,2 \ldots n) \tag{17-1}$$

$$C_p G\,\frac{dT_g}{dZ} = \sum_i \bar{k}_{s,i}\,\frac{s}{v}\,C_i \rho_g\,Q_i \tag{17-2}$$

$$- \frac{dP}{dZ} = \frac{G^2}{2\rho g_c \varepsilon^2}\,\frac{f}{L} \tag{17-3}$$

An auxiliary relationship of gas temperature and catalyst temperature is needed to allow calculation of the rate constants, $\bar{k}_{s,i}$ at the solid (catalyst) temperature:

$$T_s = T_g + \frac{1}{h}\sum_i \bar{k}_{s,i}\,C_i \rho_g Q_i \tag{17-4}$$

The other parameters appearing in these equations may be calculated from correlations:

$$\frac{1}{\bar{k}_{s,i}} = \frac{1}{k_{mt,i}} + \frac{1}{k_{chem,\,i}} \tag{17-5}$$

$$k_{chem,i} = A_i \exp\left[\left(\frac{Ea}{Rg}\right)_i \left(\frac{1}{T_{ref}} - \frac{1}{T_s}\right)\right] \tag{17-6}$$

$$h = \frac{k_g}{L} N_{Nu} \tag{17-7}$$

$$k_{mt,i} = \frac{D_i}{L} N_{Sh} \tag{17-8}$$

For the simple case in which the heating value of the fume stream is negligible, all of the Q_i's are taken as zero. In this case, $T_s = T_g$ = constant, ρ_g is constant, and Equations 17-1 and 17-3 become integrable (analytically). (Equations 17-2 and 17-4 are trivial for this case.)

For the more general case, when numerical methods must be used to solve the equation set, it is convenient to rewrite Equations 17-1 in terms of the fractional concentration (rather than actual concentration), y_i of the hydrocarbon species.

$$y_i = C_i / C^o{}_i$$

where $C^o{}_i$ is the initial concentration upstream of the catalyst bed.

$$\frac{dy_i}{dZ} = - \overline{k}_{s,i} \, y_i \, \frac{s}{v} \, \frac{\rho_g}{G} \tag{17-9}$$

or

$$\frac{d \ln y_i}{dZ} = - \overline{k}_{s,i} \, \frac{s}{v} \, \frac{\rho_g}{G} \tag{17-10}$$

The equation for gas temperature change, ΔT_g, over any interval of bed length may be written in terms of the changes in the dimensionless concentrations, Δy_i, over the same length:

$$\Delta T_g = - \frac{1}{C_p} \sum_i \Delta y_i \, C^o{}_i \, Q_i \tag{17-11}$$

For a sufficiently short interval of bed length, Equation 17-4 for the solid temperature can also be written in terms of the Δy_i over the length ΔZ:

$$\overline{T}_s \cong \overline{T}_g - \frac{G}{h \frac{s}{v} \Delta Z} \sum_i \Delta y_i \, C^o{}_i \, Q_i \tag{17-12}$$

where the temperatures, $\overline{T}_s$ and $\overline{T}_g$ are average values over the increment of length. The Equations 17-3, 17-5 to 17-8, 17-10, 17-11 and 17-12 form a convenient set for numerical solution for the general case when the Q_i are not zero. This set is reproduced in Table 17-1.

Table 17-1. RECOMMENDED EQUATION SET FOR NUMERICAL SOLUTION

$$\frac{d \ln y_i}{dZ} = -\bar{k}_{s,i} \frac{s}{v} \frac{\rho_g}{G} \quad (i = 1,2. . . n) \tag{17-10}$$

$$\Delta T_g = -\frac{1}{C_p} \sum_i \Delta y_i \, C^o_i \, Q_i \tag{17-11}$$

$$\bar{T}_s = \bar{T}_g - \frac{G}{h\frac{s}{v}\Delta Z} \sum_i \Delta y_i \, C^o_i \, Q_i \tag{17-12}$$

$$-\frac{dP}{dZ} = \frac{G^2}{2\rho_g \, g_c \varepsilon^2} \, \frac{f}{L} \tag{17-3}$$

$$\frac{1}{\bar{k}_{s,i}} = \frac{1}{k_{mt,i}} + \frac{1}{k_{chem,i}} \tag{17-5}$$

$$k_{chem,i} = A_i \exp\left[\left(\frac{Ea}{Rg}\right)_i \left(\frac{1}{T_{ref}} - \frac{1}{T_s}\right)\right] \tag{17-6}$$

$$h = \frac{k_g}{L} \, N_{Nu} \tag{17-7}$$

$$k_{mt,i} = \frac{D_i}{L} N_{Sh} \tag{17-8}$$

Tables 17-2 and 17-3 follow

250h

TABLE 17-2

```
C         LOGICAL SECTION FROM FORTRAN PROGRAM FOR SOLUTION OF DESIGN
C                     EQUATIONS FOR CATALYTIC AFTERBURNERS

C     DIVIDE TOTAL BED LENGTH FOR CALCULATION, ZTOTAL, INTO N SEGMENTS
      DELZ = ZTOTAL/FLOAT(NSEG)
C     SET INITIAL CONDITIONS
C     INDEX(I) IDENTIFIES HYDROCARBON SPECIES
C     INDEX(N) IDENTIFIES LENGTH SEGMENT IN BED
      P(1) = PIN
C     N=1 IS INLET TO BED, N=NSEG+1 IS OUTLET
C     N=N IS INLET TO NTH SEGMENT, N=N+1 IS OUTLET FROM NTH SEGMENT
      T(1) = TIN
C     T IS THE GAS TEMPERATURE
      DO 1 I=1,NHC
      Y(I,1) = 1.0
C     Y(I) IS THE FRACTION REMAINING OF THE ORIGINAL CONCENTRATION,
C      CIN(I), OF THE I'TH SPECIES
C     CALCULATE CIN,Q PRODUCT.  Q IS THE HEAT OF COMBUSTION OF SPECIES
    1 CQ(I) = CIN(I)*Q(I)
C     START CALCULATION FOR LENGTH SEGMENTS
      DO 2 N=1,NSEG
      TBAR = T(N)
C     TBAR IS ARITH. AVG. TEMP. OF GAS ACROSS SEGMENT
      TS = T(N)
C     TS IS MEAN TEMP OF CATALYST SURFACE IN SEGMENT
    4 TSTOR = TS
C     TSTOR IS MOST RECENT ASSUMED CATALYST TEMP FOR SEGMENT
      CALL RATE
C     RATE SUBROUTINE USES CORRELATIONS TO CALCULATE H(HT COEFF),
C      F(FRICTION FACT), KSBAR(I) (OVERALL 1ST ORDER RATE CONST)  AND
C      RHO(GAS DENSITY)
      DELT = 0.0
C     DELT IS THE GAS TEMP RISE ACROSS INCREMENT
      DO 3 I=1,NHC
      Y(N+1,I) = Y(N,1)/EXP(KSBAR(I)*SONV*DELZ*RHO/G)
C     PSEUDOANALYTICAL SOL'N USED FOR EQN 17-10
C     SONV IS SURFACE TO VOLUME RATIO OF MATRIX, G IS GAS MASS VELOCITY
    3 DELT = DELT + CQ(I)*(Y(N,I)-Y(N+1,I))
      T(N+1) = T(N) + DELT/CP
      TBAR = (T(N+1) + T(N))/2.0
      TS = TBAR + DELT*G/(H*SONV*DELZ)
C     REPEAT CALCULATION UNTIL CATALYST TEMP, TS, CONVERGES WITHIN 1DEGF
      IF(ABS(TS-TSTOR).GT.1.0) GO TO 4
      P(N+1) = P(N) - (G**2)*F*DELZ/RHO/64.4/(EPSILN**2)/L
C     EPSILN IS VOID FRACTION, L, IS CHARACTERISTIC LENGTH FOR MATRIX
C     CALCULATIONS COMPLETED FOR NTH SEGMENT. PROCEDE TO NEXT SEGMENT
C     INTERMEDIATE PRINTOUT INSERTED HERE
    2 CONTINUE
C     FINAL OUTPUT FOR CASE CALCULATIONS
```

Table 17-3. GEOMETRIC PROPERTIES OF AFTERBURNER CATALYSTS AND RECOMMENDED CORRELATIONS FOR TRANSPORT PROPERTIES

Catalyst Type	S/V (ft^{-1})	ε	L (ft)	X (ft)	N_{Nu}, N_{Sh}, f
Torvex 2B (1/8 in. hexcell honeycomb)	268	0.61	0.0091	0.0833	$N_{Nu} = 3.66\,(1 + 0.095\,N_{Re}\,N_{Pr}\,\frac{L}{X})^{0.45}$ $N_{Sh} = 3.66\,(1 + 0.095\,N_{Re}\,N_{Sc}\,\frac{L}{X})^{0.45}$ $f = 64/N_{Re}\,(1 + 0.0445\,N_{Re}\,\frac{L}{X})^{0.5}$
Thermocomb 8 c/in. honeycomb	695	0.6	0.00345	normally 0.167 - 0.333	$N_{Nu} = 2.35\,(1 + 0.095\,N_{Re}\,N_{Pr}\,\frac{L}{X})^{0.45}$ $N_{Sh} = 2.35\,(1 + 0.095\,N_{Re}\,N_{Sc}\,\frac{L}{X})^{0.45}$ $f = 53.3/N_{Re}\,(1 + 0.0445\,N_{Re}\,\frac{L}{X})^{0.5}$
Oxycats	36.5	0.515	0.0133	-	$N_{Nu} = 0.51\,N_{Re}^{0.56}\,N_{Pr}^{0.333}$ $N_{Sh} = 0.51\,N_{Re}^{0.56}\,N_{Sc}^{0.333}$ $f = 0.5/N_{Re}^{0.15}$
Metal ribbon "D" series	336	0.93	0.004	-	$N_{Nu} = 0.55\,N_{Re}^{0.5}\,N_{Pr}^{0.333}$ $N_{Sh} = 0.55\,N_{Re}^{0.5}\,N_{Sc}^{0.333}$ $f = (1-\varepsilon)/\varepsilon(5.2 + 3600\,(1-\varepsilon)/N_{Re})$ $-\Delta P = \int_0^z G^2/2\varepsilon^2\rho_g g_c f/L\,dZ + 2.1\,G^2/2\rho_g\,g_c$
Spherical catalyst pellets	$\frac{6\varepsilon}{dp}$	normally 0.35 - 0.4	dp	-	$N_{Nu} = 2.42\,N_{Re}^{1/3}\,N_{Pr}^{1/3} + 0.129\,N_{Re}^{0.8}\,P_r^{0.4} + 1.4\,N_{Re}^{0.2}$ $N_{Sh} = 2.42\,N_{Re}^{1/3}\,N_{Sc}^{1/2} + 0.129\,N_{Re}^{0.8}\,N_{Sc}^{0.4} + 0.4\,N_{Re}^{0.2}$ $f = (1-\varepsilon)/\,\varepsilon(3.5 + 300\,(1-\varepsilon)/N_{Re})$

2500

A short section presenting the logic for numerical solution of the set is presented as a partial FORTRAN computer program in Table 17-2. A complete program using this logic has been used in preparing the figures presented in Chapter 4 illustrating the effects of operating variables on catalyst performance. No problems with convergence or stability of the integration procedure were encountered when using ten axial increments through the bed in the numerical integrations, even under conditions where rates were high. Other numerical methods could be used. However, the pseudo-analytical solution of Equation 17-10 is recommended since it offers a strong stabilizing influence on the numerical solution when source terms are large.

17.3 Recommended Correlations of Rate Parameters for Afterburner Catalyst Matrices

17.3.1 Transport Coefficients

Table 17-3 presents geometric properties and recommended correlations for Nusselt and Sherwood numbers and friction factors for the four matrix types in commercial use, and for spherical catalyst pellets. Geometrical properties are based on data from catalyst manufacturers. The basis of Nusselt and Sherwood number correlations was discussed in Chapter 16.

Friction factor correlations for the two honeycomb supports are based on curve fitting numerical solutions for "$f_{apparent}$" presented in Figure 6-22 of Reference 16-2. The tube cross section is assumed to be circular for the "TORVEX" and triangular for the "THERMACOMB" material. These correlations for friction factor have also been confirmed by comparison with experimental data covering a wide range of flow rates and temperatures for both "TORVEX" and "THERMACOMB" catalysts. Data were made available for this comparison by the Industrial and Biochemicals Department, E. I. duPont.

The "OXYCAT" configuration is a close approach to a conventional heat exchanger tube bundle. The friction factor correlation recommended is taken directly from Reference 16-4 for cross flow through tube bundles. No confirmatory data for the correlation are available.

The friction factor expression for the metal ribbon catalyst is based on data presented in Reference 4-5 and other data from the manufacturer. The correlation is in the conventional Ergun form, but coefficients are unusually high. Because this catalyst is normally used in thin beds and with low overall pressure drop, the inlet and outlet retaining screens contribute significantly to the overall pressure drop through the element. An expression for ΔP incorporating an additional term which accounts for this contribution is included in Table 17-3.

All of the transport property correlations for spherical catalyst pellets are those recommended by Beek[16-11].

17.3.2 Chemical Rate Constants

The publicly available data on afterburner catalyst performance provide a meager basis for "chemical" rate constant correlations. A few estimates of rate parameters are presented in Table 17-4. These are based on normal design specifications, published "ignition" and operating temperatures and the limited performance data available in the literature. They provide some guide as to the values to be expected for these parameters. It is to be hoped that catalyst and afterburner manufacturers can supply more and better data by analysis of their performance experience.

Table 17-4. ESTIMATED "CHEMICAL RATE" PARAMETERS FOR AFTERBURNER CATALYSTS

$$k_{chem,i} = A_i \ \exp\left[\left(\frac{Ea}{R_g}\right)_i \left(\frac{1}{T_{ref}} - \frac{1}{T_s}\right)\right]$$

Component	$k_{chem,i}$ at T_{Ref} a) (ft/sec)	Ea (Kcal/gm mole)	Ea/Rg (°R)
Alumina-Based Catalysts			
H_2	8.0	8	7,250
CO	0.45	11	10,000
Solvent HC's	0.1 - 0.2	11	10,000
CH_4	0.005	15	13,600
Metal-Based Catalysts			
Toluene/xylene	0.2 - 0.3	15	13,600
MEK	0.03	15	13,600
n-heptane	0.005	15	13,600

a) For all of the cases presented in the Table, T_{Ref} = 1010°R (550°F).

17.4 Estimation of Performance with Partially Deactivated Catalyst

An estimate of the consequences of partial deactivation of the catalyst can be obtained if the calculation made for fresh catalyst is repeated using values of $k_{s,i}$ modified to account for deactivation effects.

For deactivation by aging or poisoning in which active sites throughout the substrate are reduced in number, the values of $k_{s,i}$ could be calculated using a value of $k_{chem,i}$ defined.

$$k_{chem,i} = f_1 A_i \exp\left[\left(\frac{Ea}{Rg}\right)_i \left(\frac{1}{T_{ref}} - \frac{1}{T_s}\right)\right]$$

where f_1 is a multiplier between 0 and 1. The quantity, $1 - f_1$ can be regarded as the fractional deactivation of the catalyst sites.

If activity decline is due to surface coating by particulates, the chemical activity would be unchanged. However, mass transfer resistance would be increased since reactants would have to diffuse through the inert layer to reach the active catalyst. For this case, the mass transfer coefficient, $k_{mt,i}$ might be defined:

$$k_{mt,i} = \frac{1}{\frac{L}{D_i N_{Sh}} + \frac{1}{k_f}}$$

where $1/k_f$ is the effective diffusive resistance of the surface coating layer.

For more complex patterns of deactivation, adaptations of the calculation method could be developed.

NOMENCLATURE

(Chapter 17)

A_i	Pre-exponential factor, value of $k_{chem,i}$ at reference temperature (ft/sec).
C_p	Heat capacity of gas stream (Btu/lb-°F)
C_i	Concentration of ith species in waste stream (lb mole/lb waste stream)
$C_i{}^o$	Concentration of ith species at catalyst bed inlet end (lb mole/lb)
d_p	Particle diameter of spherical catalyst pellets (ft)
D_e	Equivalent diameter of catalyst support (ft)
$(Ea/Rg)_i$	Exponential factor for temperature dependence of chemical rate ($°R^{-1}$)
f	Friction factor
G	Superficial mass velocity of gas through matrix (lb/sec-ft^2)
g_c	Gravitational conversion factor (lbf-ft/lbm-sec^2)
h	Gas-to-solid heat transfer coefficient (Btu/sec-ft -°F)
k_g	Thermal conductivity of gas (Btu/sec-ft-°F)
$k_{chem,i}$	Rate constant for oxidation of ith species (ft/sec)
$k_{mt,i}$	Rate constant for mass transfer of ith species (ft/sec)
$\overline{k}_{s,i}$	Overall rate constant for oxidation of ith species (ft/sec)
L	Matrix length dimension used in dimensionless numbers and in pressure drop equation (ft)
N_{Nu}	Dimensionless Nusselt number, hL/k_g
N_{Pr}	Dimensionless Prandtl number, $C_p\mu/k_g$
N_{Re}	Dimensionless Reynolds number, $LG/\mu\varepsilon$
N_{Sc}	Dimensionless Schmidt number, $\mu/\rho_g D_i$
N_{Sh}	Dimensionless Sherwood number, $k_{mt,i}\ L/D_i$
P, ΔP	Pressure, pressure drop through matrix (lbf/ft^2)

NOMENCLATURE (CONT)

(Chapter 17)

Q_i	Heat of oxidation of ith species (Btu/lb mole)
r_h	Hydraulic radius of honeycomb matrix (ft)
s/v	Surface to volume ratio of matrix (ft^{-1})
T_g	Gas temperature (°F or °R)
T_{ref}	Reference temperature for chemical rate constant (°F or °R)
T_s	Temperature of catalyst surface (°F or °R)
X	Length between boundary layer interruptions in honeycomb matrices (thickness of individual layers of catalyst in the bed) (ft)
y_i	Dimensionless (fractional) concentration of ith species, C_i/C_i°
Z	Length dimension in direction of flow through the catalyst bed (ft)
ε	Fraction voids of matrix
ρg	Gas density (lb/ft^3)
μ	Gas viscosity (lb/ft-sec)

Chapter 18. HEAT RECOVERY - DESIGN FUNDAMENTALS

18.1 Introduction

Chapter 6 described various means for achieving heat recovery in a system to be equipped with an afterburner. This section gives a presentation of working relations and data intended for planning the installation, for checking design proposals, and for evaluating performance of the heat recovery equipment, particularly of cross-flow tubular heat exchangers. The section is not meant to be complete enough to give a working basis for design of equipment. However, an engineer having a general familiarity with heat transfer apparatus, with fluid flow, and with elementary structural analysis, could undoubtedly prepare a design. References will be included, where appropriate, for more complete data or more specialized calculations.

The heat transfer relations included here are:

Quantity of Heat - which may be recovered from cooling a hot stream, or or which is required to heat a cool stream (Equations 18-1 through 18-4, Section 18.2.1)

Heat Transfer Rate Equations - (Equation 18- 5, Section 18.2.2)

Heat Transfer Coefficients - (Figures18-2,-3, Section 18.2.3)

Heat Exchanger Effectiveness - (Equation 18- 8, Section 18.2.4)

Number of Transfer Units - (Equation 18- 9, Section 18.2.5)

Capacity Rate Ratio - (Section 18.2.6)

Cross-Flow Exchanger Performance - (Section 18.2.7)

Exchanger Pressure Drop - (Section 18.3)

Miscellaneous Equations and Relations - (Section 18.3.6)

18.2 Equations and Relations for Heat Transfer

18.2.1 Quantity of Heat to be Transferred

$$Q = 60w\ Cp\ (T_{final} - T_{initial}) \qquad (18\text{-}1)$$

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where Q is the amount of heat, Btu/hr, for producing the change of temperature between the limits, $T_{initial}$ and T_{final}. It is assumed there are no condensing or evaporating liquids (that is, there are no phase changes).

w is the mass flow rate of gases involved, lb/min.

Cp is the average value of specific heat over the temperature range, $(T_{final} - T_{initial})$ at constant pressure, in Btu/(lb-°F). The specific heat of a gas varies in a non-linear way with temperature, but may be approximated for the purposes here by using the average "instantaneous" values between the two limiting temperatures.

$$(Cp)_{Average} = \frac{(Cp)_{T_{initial}} + (Cp)_{T_{final}}}{2} \qquad (18\text{-}2)$$

T is temperature in degrees F.

Values of specific heat for air and different gases may be taken from Figure 18-1.

The specific heat of a mixture of gases is found from the specific heats of the components, all taken over the same temperature range,

$Cp_1, Cp_2 \ldots Cp_n$

$$Cp = w_1Cp_1 + w_2Cp_2 \ldots w_nCp_n \qquad (18\text{-}3)$$

where $w_1, w_2 \ldots w_n$ are the weight fractions of the n components. If the composition is given in terms of volume fractions, $v_1, v_2, \ldots v_n$, then the corresponding weight fractions can be found by

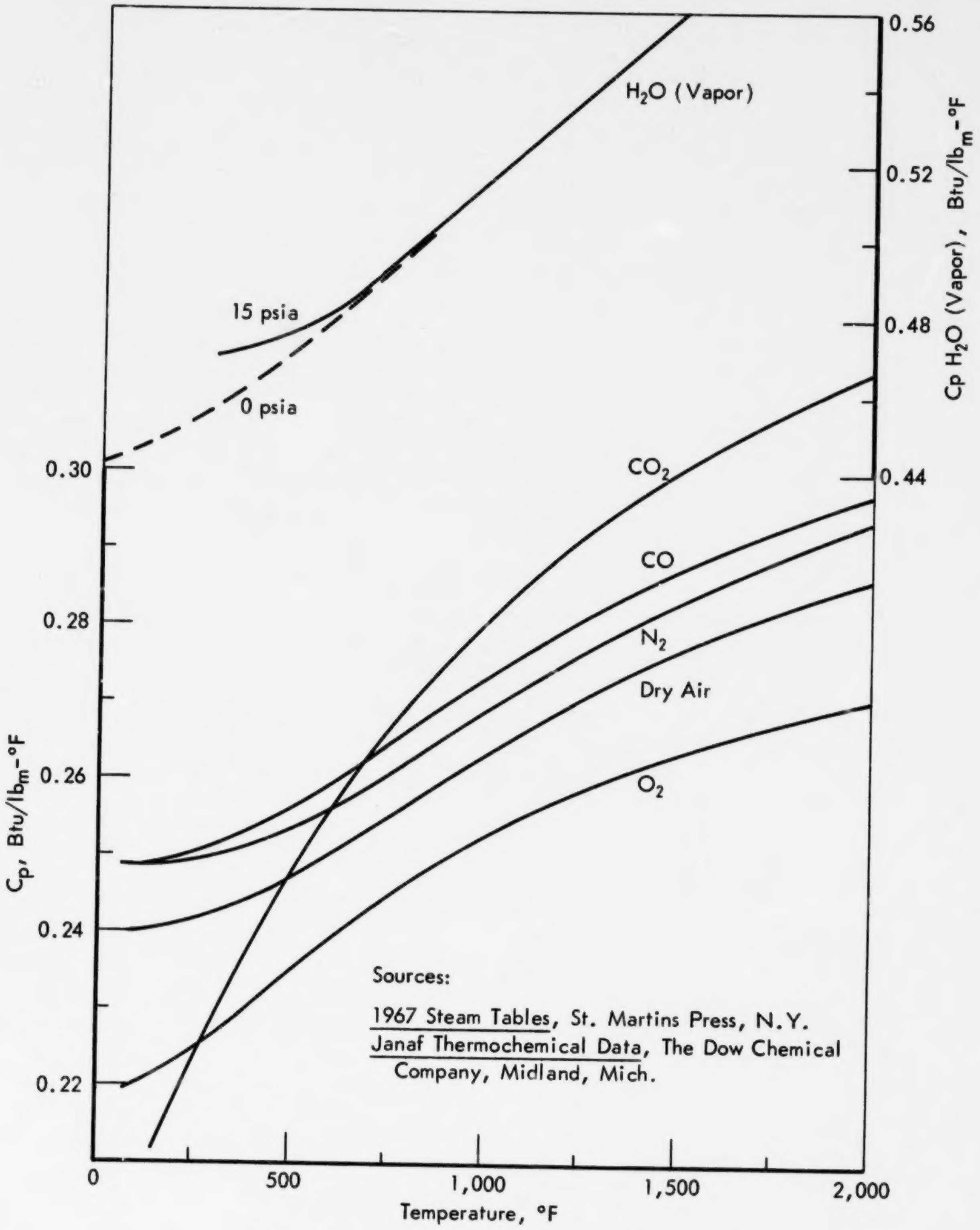
$$w_1 = \frac{v_1M_1}{v_1M_1 + v_2M_2 + \ldots v_nM_n}$$

$$w_2 = \frac{v_2M_2}{v_1M_1 + v_2M_2 + \ldots v_nM_n} \qquad (18\text{-}4)$$

where $M_1, M_2 \ldots$ are the molecular weights of the gases, as given in Table 18-1. (Note: mole-fraction is the same as volume fraction.)

In this simplified approach, all gases are treated as "perfect gases", and all vapor components are assumed to have a low partial pressure so that perfect gas behavior is approximated.

Figure 18-1 follows



Sources:

1967 Steam Tables, St. Martins Press, N.Y.
Janaf Thermochemical Data, The Dow Chemical Company, Midland, Mich.

Figure 18-1. SPECIFIC HEATS OF GASES

Table 18-1. MOLECULAR WEIGHTS OF GASES

Dry Air*		29.0
Flue Gas	(Nat. Gas, 10% Excess Air)	29.3
	(Nat. Gas, 400% " ")	29.0
O_2		32.00
N_2		28.02
CO		28.01
CO_2		44.01
H_2O		18.02
CH_4 Methane		16.04
C_2H_6 Ethane		30.07
C_2H_4 Ethylene		28.05
C_3H_8 Propane		44.09
C_3H_6 Propylene		42.08

*Composition O_2	20.91% vol	23.15%wt	
N_2, A	79.09	76.85	
	100.00	100.00	0

18.2.2 Heat Transfer Rate Equations

The heat transfer rate between two fluids separated by a wall, or between a fluid and a wall, is calculated by the equation

$$q = uA\,(T_h - T_c) \qquad (18\text{-}5)$$

where q = heat transfer rate, Btu/hr
A = heat transfer area, ft^2
T_h = hot fluid (or surface) temperature, °F
T_c = cold fluid (or surface) temperature, °F
u = overall transfer coefficient Btu/(hr-ft^2 - deg F)

18.2.3 Heat Transfer Coefficients

The overall transfer coefficient is made up of a number of components, depending on the nature of the heat transfer system.

18.2.3.1 Film Coefficients, h_f

For a gas-to-gas exhcnager there will be two film coefficients; one on the inside, one on the outside of the tube. h_f accounts for the insulating effect caused by a thin layer of slowly moving gas, adjacent to the metal surface of the tube. h_f is a complex function of surface geometry, flow channel geometry, fluid properties (density, viscosity) and average fluid velocity over the surface. Values of h_f have been plotted for a useful range of gas velocities, for flow of air or flue gas inside tubes (Figure 18-2) and across tubes (Figure 18-3). The choice of the correct velocity to be used in these graphs will be discussed later. For other geometries, and for a more complete treatment of a complex subject, consult W. M. Kays, A. L. London, Compact Heat Exchangers, Reference 18-1.

18.2.3.2 Wall Conductance, C_w

For thin-walled tubing this may be approximated by

$$C_w = \frac{t}{12k} \qquad (18\text{-}6)$$

where t is the thickness of the tube wall in inches, k is the thermal conductivity of the conducting metal in Btu/(hr-ft^2-deg F/ft). Values of k for common metals are given in Figure 18-4. k is evaluated at approximately the median temperature between flue gas and fume stream. For most afterburner applications this term may be neglected.

18.2.3.3 Fouling Factor, f

Thick fouling layers on the tube walls may predominate in resisting heat transfer from one gas to the other. However, predicting f in the absence of experience on a particular application is largely arbitrary. Conditions may range from clean (f = 0) to extremely dirty. Figure 18-5 gives values of f that would correspond to a thickness of asphalt and to petroleum coke. At low

Figures 18-2 through 18-5 follow

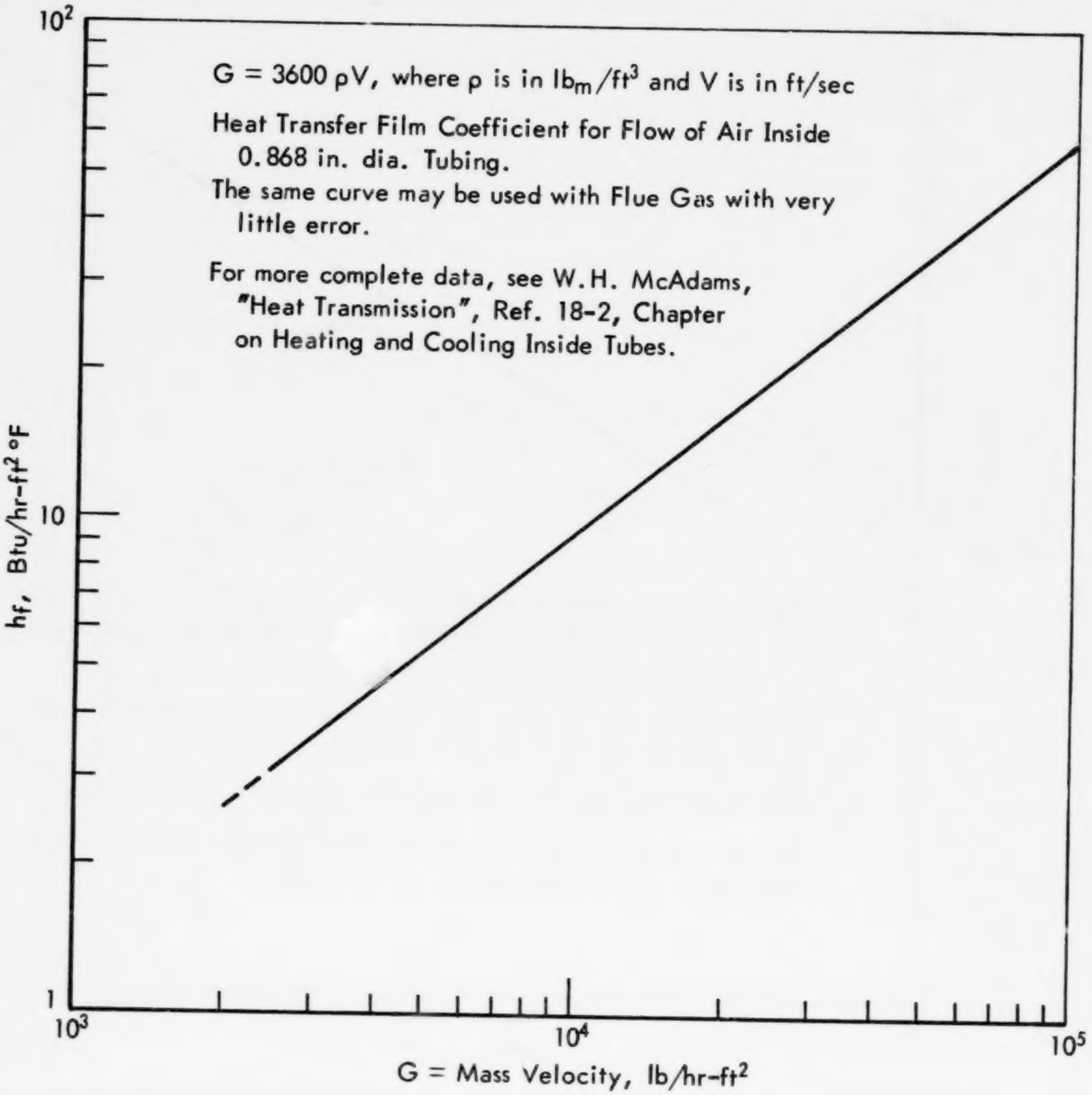


Figure 18-2. HEAT TRANSFER FILM COEFFICIENT FOR FLOW OF AIR INSIDE TUBES

260 h

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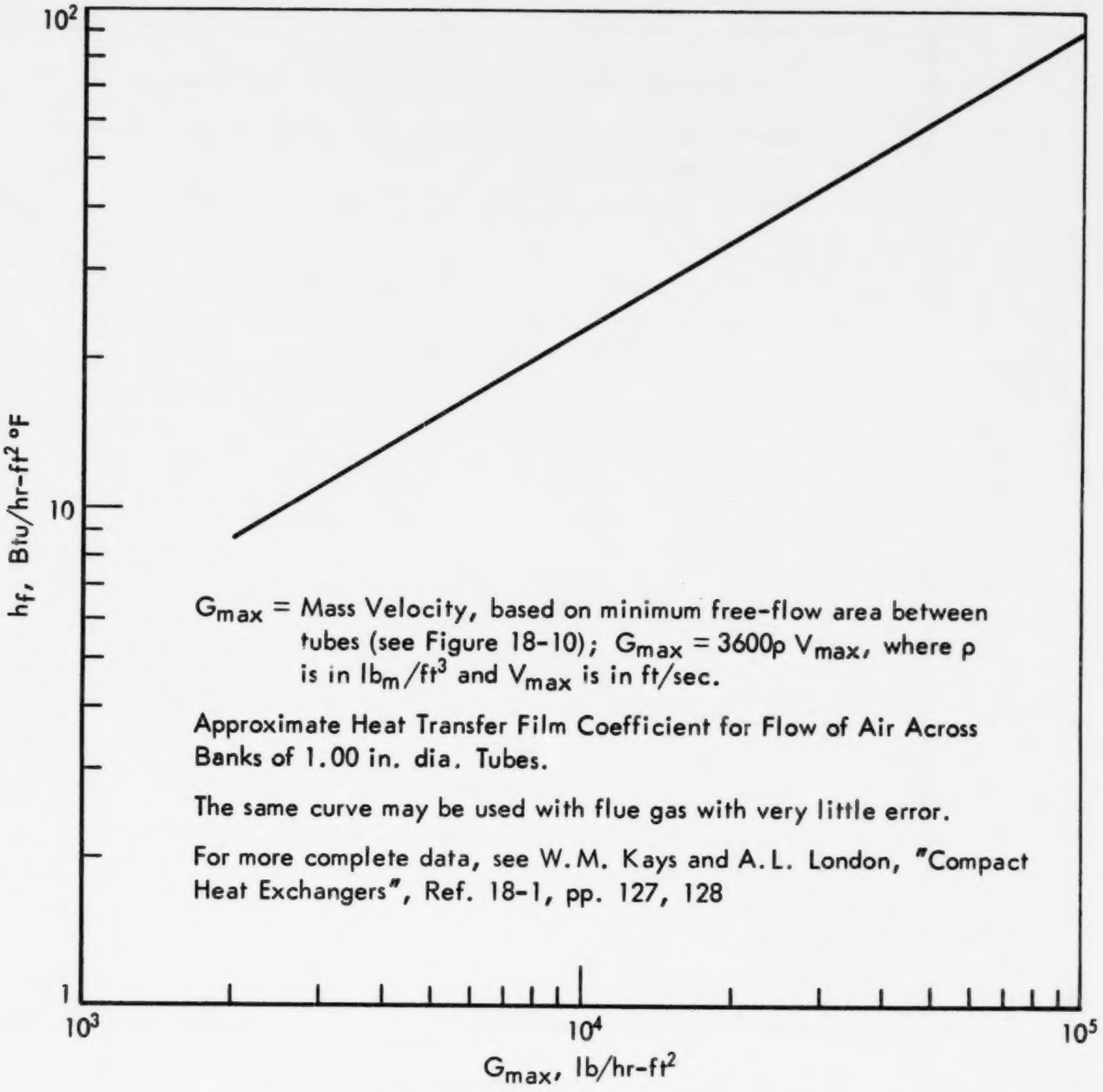

G_{max} = Mass Velocity, based on minimum free-flow area between tubes (see Figure 18-10); $G_{max} = 3600\rho\, V_{max}$, where ρ is in lb_m/ft^3 and V_{max} is in ft/sec.

Approximate Heat Transfer Film Coefficient for Flow of Air Across Banks of 1.00 in. dia. Tubes.

The same curve may be used with flue gas with very little error.

For more complete data, see W. M. Kays and A. L. London, "Compact Heat Exchangers", Ref. 18-1, pp. 127, 128

Figure 18-3. HEAT TRANSFER FILM COEFFICIENT FOR FLOW OF AIR ACROSS BANKS OF TUBES

260c

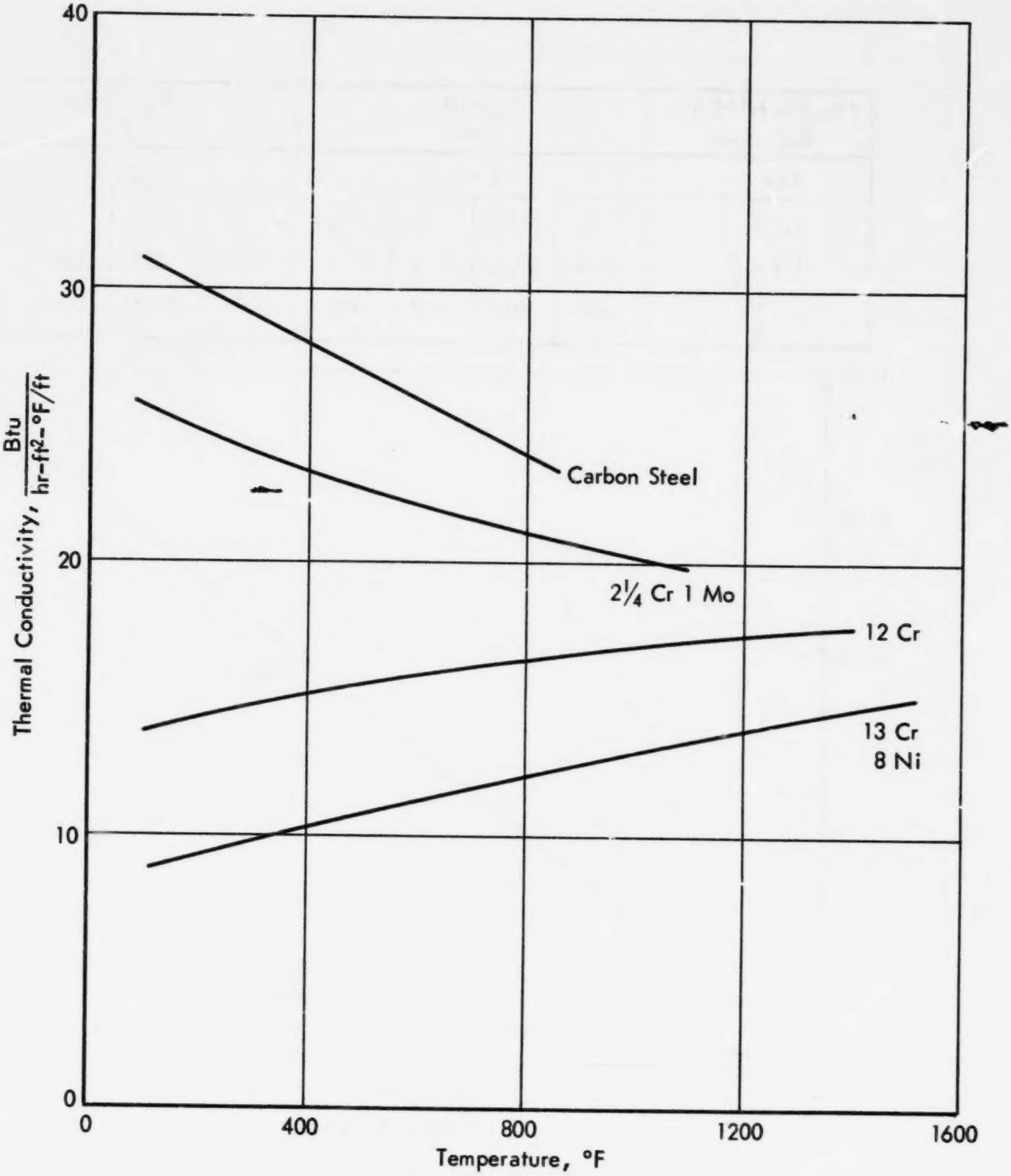

Figure 18-4. THERMAL CONDUCTIVITY OF METALS

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Btu/(hr ft^2 °F/ft) Ref. 18-2	Asphalt .43				Pet. Coke 3.4			
X12	5.16				40.8			
ℓ, in.	1/4	1/16	1/8	3/16	1/4	16	1/8	3/16
$1/f = \frac{\ell}{k}$	20.4	82.6	41.3	27.5	167	653	326	218
$f = \frac{k}{\ell}$	.049	.012	.024	.036	.006	.0015	.003	.0046

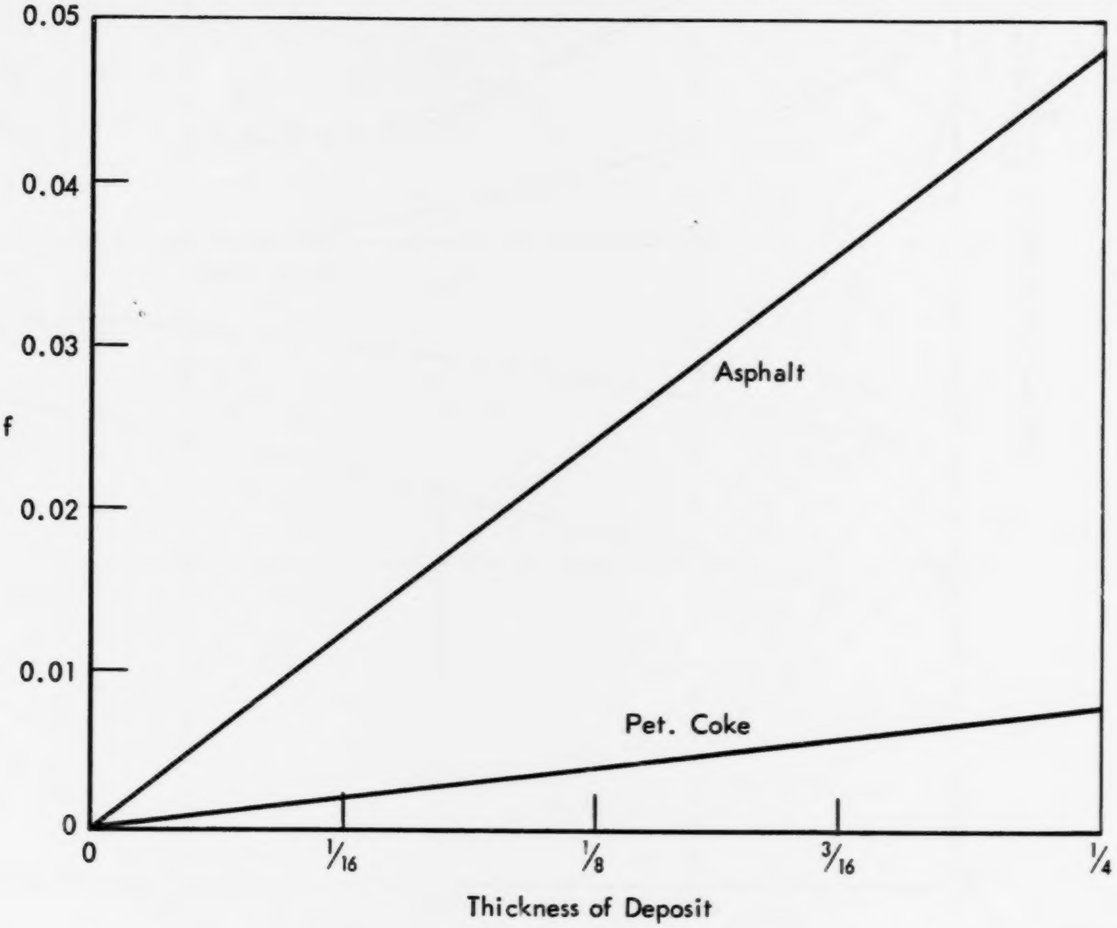

Figure 18-5. FOULING FACTOR

2602

temperatures, greasy materials, polymerizing resins, etc., will be more or less like asphalt. At elevated temperature (say 500°F or more), any such materials are likely to turn into coke-like layers; hence, coke is probably more typical of what would be encountered in an exchanger exposed to 1500°F flue gas on the other side of the tube wall. Fouling material may not deposit uniformly over the exchanger surface, and the conversion to coke will be affected by the local tube wall temperature in the exchanger. Where possible, the nature of fouling in a given service should be established by experience.

Table 18-2 gives suggested fouling factors for specific services.

Table 18-2. SUGGESTED FOULING FACTORS FOR RECUPERATIVE HEAT EXCHANGERS

Service	Factor
Deodorizers, odor control	.005
Paint drying ovens	.01
Lacquer drying ovens	.005
Print drying ovens	.005
Steel drum reconditioning	.02
Meat packing and rendering	.05
Pattern Curing Ovens (Foundry)	.01
Biological Cleanings of Contaminated Air	.005

Note: These factors are highly tentative and should be modified as experience warrants.

18.2.3.4 Overall Coefficient

The overall transfer coefficient is calculated,

$$\frac{1}{u} = \frac{1}{(h_f)_{\text{inside tube}}} + f + C_w + \frac{d_o}{d_i} \frac{1}{h_{f_{\text{outside tube}}}} \tag{18-7}$$

This is evaluated on the basis of the inside area of the tube (d_i = inside diameter, d_o = outside diameter), with the further supposition that the cool fume stream will pass inside the tubes.

18.2.4 Heat Exchange Effectiveness, E

The effectiveness is defined (Ref. 18-1):

$$E = \frac{Q}{Q_{max}} = \frac{\overline{C}_{fume}\left(T_{fume\ leaving} - T_{fume\ entering}\right)}{\overline{C}_{min}\left(T_{flue\ gas\ entering} - T_{fume\ entering}\right)} \qquad (18\text{-}8A)$$

$$\text{Also} = \frac{\overline{C}_{flue\ gas}}{\overline{C}_{min}} \frac{\left(T_{flue\ gas\ entering} - T_{flue\ gas\ leaving}\right)}{\left(T_{flue\ gas\ entering} - T_{fume\ entering}\right)} \qquad (18\text{-}8B)$$

$\overline{C}$ = wC_p for the appropriate stream; it is the heating capacity rate for the stream

$\overline{C}_{min}$ = Refers to that stream having the lowest heating capacity rate. Since the fume stream has less mass flow than the flue gas (which includes the fuel and added combustion air), and C_p fume ≈ C_p flue gas

$$\overline{C}_{min} = \overline{C}_{fume} = (w_{fume}\ C_{Pfume})$$

C_p = The average specific heat of the gas at constant pressure over the temperature range Btu/(lb-deg F). (See Eq. 18-3)

w = Mass flow rate of stream, lb/hr.

Equation 18-8A thus reduced to:

$$E = \frac{\left(T_{fume\ leaving} - T_{fume\ entering}\right)}{\left(T_{flue\ gas\ entering} - T_{fume\ entering}\right)} \qquad (18\text{-}8C)$$

Equation 18-8B becomes:

$$E = \frac{\overline{C}_{flue\ gas}}{\overline{C}_{fume}} \frac{\left(T_{flue\ gas\ entering} - T_{flue\ gas\ leaving}\right)}{\left(T_{flue\ gas\ entering} - T_{fume\ entering}\right)} \qquad (18\text{-}8D)$$

The effectiveness is a measure of the performance of the exchanger, comparing the actual recovery of heat with that which might be obtained using an ideal exchanger of infinite size.

18.2.5 Number of Transfer Units, N_{tu}

The number of transfer units is a non-dimensional grouping of terms that characterizes the size of the exchanger.

$$N_{tu} = \frac{Au}{(w\ C_p)} \quad ; \quad A = \frac{w\ C_p\ N_{tu}}{u} \qquad (18\text{-}9)$$

where A is the heat transfer area of the exchanger inside the tubes, ft^2.

$$A = (\pi/4)\ (1/144)\ NL\ d_i^2$$
$$= .00545\ NL\ d_i^2 \qquad (18\text{-}10)$$

taking A = number of tubes
L = length of tubes between tube plates, ft.
d_i = tube inside diameter, in.

The exposed area of the tube sheets is ignored. u is the overall heat transfer coefficient, Btu/(hr-ft^2-deg F) based on inside area, w is the fume gas flow rate, lb/hr, and C_p is the average specific heat of the fume stream over the temperature interval.

18.2.6 Capacity Rate Ratio

$$R = \frac{C_{fume}}{C_{flue\ gas}} = \frac{w_{fume\ stream} \times C_{p_{fume\ stream}}}{w_{flue\ gas\ stream} \times C_{p_{flue\ gas\ stream}}} \qquad (18\text{-}11)$$

Again, w is in lb/hr, and C_p is in Btu/(lb-deg F) and is taken at the average value over the temperature interval of fume or flue gas stream. Since the flue gas stream is enlarged by the mass rate of the fuel, and the value of C_p for the flue gas is increased somewhat by the average temperature, R for this application is usually a number slightly less than 1.0.

18.2.7 Cross Flow Exchanger Performance

The heat transfer effectiveness can be expressed as a function of N_{tu}, R and flow arrangement. Solution of the mathematical equations for heat transfer effectiveness with different arrangements has been carried out by computer and is given in Figure 18-6 for a single pass cross-flow exchanger, with fume stream passing through the tubes, and flue gases across the outside of the tubes (unmixed). Figure 18-7 gives a similar plot for a 2-or-more passes, with R = 1. For more than two passes, the performance closely approaches that of a true counterflow exchanger. For a more complete exposition of the different exchanger arrangements and their effectiveness, see Kays and London, Ref. 18-1, Chapter 2.

The effectiveness-N_{tu} approach used here is preferred over the somewhat more familiar relation based on log-mean-temperature-difference, since the latter requires a solution by successive approximations for many practical problems, especially in predicting the performance of an existing exchanger, for which the geometry is fixed. It is also apparent from Figures 18.6 and 18.7 that the achievement of high effectiveness is gained only at the expense of relatively large size exchangers (high N_{tu}); also that the maximum attainable effectiveness, even for a very large size exchanger, is limited to about 60% for a single-pass cross-flow arrangement.

18.3 Exchanger Pressure Drop Calculations

The calculation of pressure drop through the heat recovery system requires some knowledge of the detailed geometry of both the exchanger and its ducting. Since the latter is subject to the great variety of the many possible installations, the notes given here relate primarily to the heat exchanger, which is assumed to be a shell-and-tube cross flow unit.

For method and data on calculating pressure losses in ducting, including elbows, bends, tees, changes in cross section, headers, etc., refer to the American Society of Heating, Refrigeration, and Air Conditioning Engineers, Handbook of Fundamentals (commonly known as the ASHRACE Guide), Reference 18-3.

The pressure drop through the exchanger is made up of several parts:

Headers (Section 18.3.1)

Entrance losses into exchanger tubes, cold side (Section 18.3.2)

Fluid acceleration from changes in temperature. Since velocities are generally low, this component of pressure drop will be ignored here.

Tube wall friction (or channel flow friction) (Section 18.3.3)

Exit pressure gain on leaving exchanger tubes, cold side (Section 18.3.2)

These pressure drops must be repeated for cross flow over tubes, hot side (Section 18.3.4)

All these items contribute to the overall pressure loss and affect the part of blower power which is chargeable to the heat recovery system.

18.3.1 Headers

The distribution of fume and flue gas streams into the exchanger is important; improper header design will lead to non-uniform gas velocities through the exchanger, possibly to impingement of fouling materials, and consequently to hot spots, warpage (or failure), and reduced effectiveness. Abrupt expansion in area from the supply duct, immediately following an elbow, for example, can give a very poor flow distribution into the exchanger. On the other hand, turning vanes in the elbow intended to improve distribution might become badly fouled and thus affect distribution--probably systems subject to such severe fouling should not be candidates for a heat exchanger. The ASHRACE Handbook (Ref. 18-3) can give some guidance on distribution and header design. Figure 18-8 gives examples. "A" shows an inclined duct arrangement that gives good distribution without turning or splitter vanes. "B" shows the use of vanes on the inlet sides of the exchanger. A relatively abrupt transition may be used on the outlet side without appreciable effect on distribution in the exchanger.

Header pressure losses, if accounted for, should be included in the ducting calculations.

18.3.2 Entrance and Exit Losses in Tubes

The entrance pressure loss and exit pressure gain for flow through the exchanger tubes are ordinarily not large, and may be neglected for most cases.

Figures 18-6 through 18-9 follow

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COURTESY: *W.M. Kays and A.L. London, "Compact Heat Exchangers", 2nd Ed., McGraw Hill, 1964, with the Permission of the Publisher.*

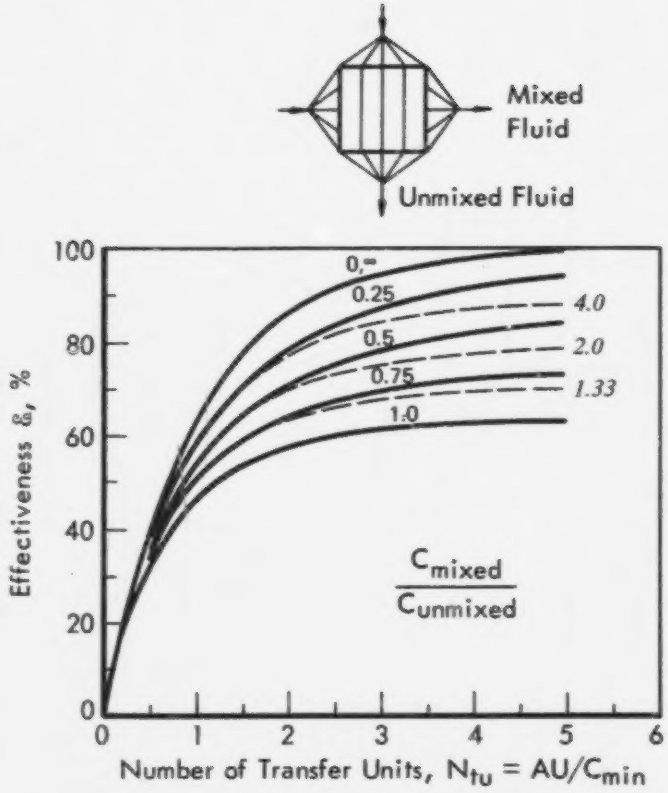


Figure 18-6. HEAT TRANSFER EFFECTIVENESS AS A FUNCTION OF NUMBER OF TRANSFER UNITS AND CAPACITY RATE RATIO; CROSSFLOW EXCHANGER WITH ONE FLUID MIXED

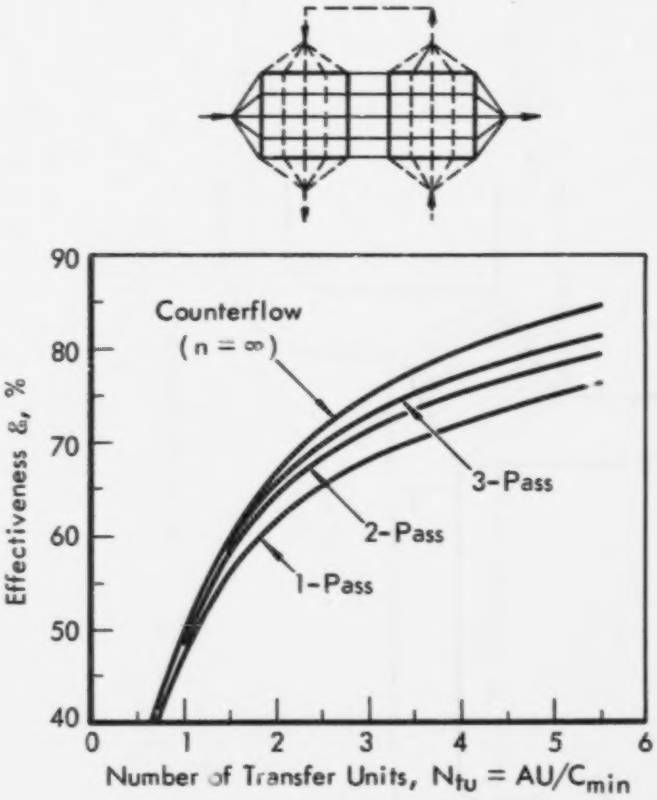
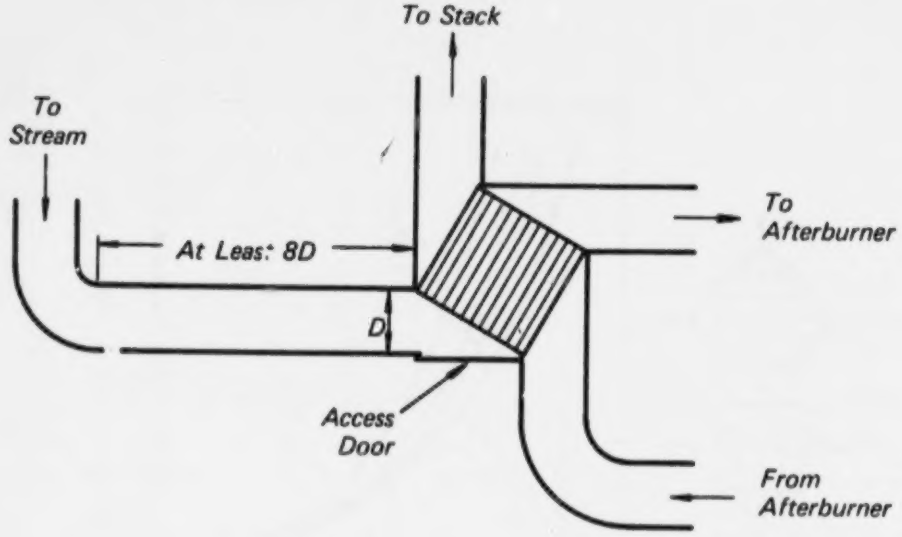
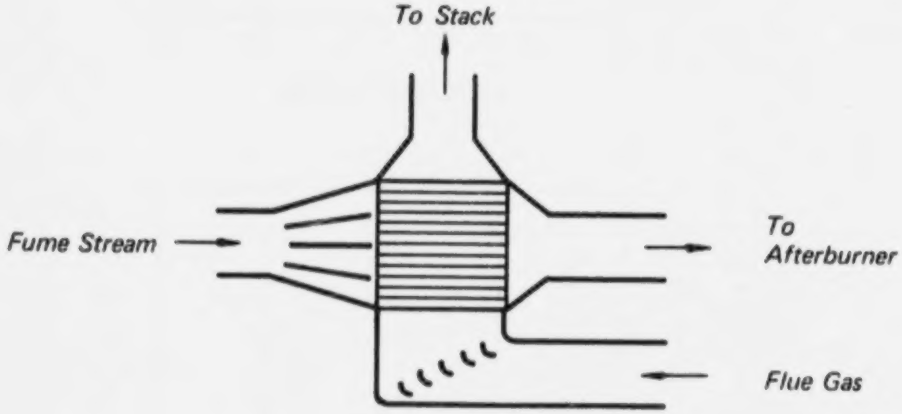


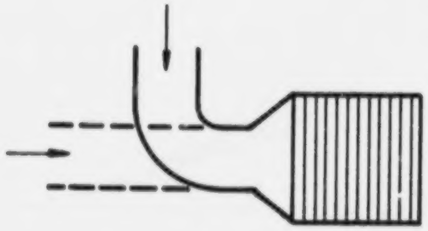
Figure 18-7. HEAT TRANSFER EFFECTIVENESS AS A FUNCTION OF NUMBER OF TRANSFER UNITS AND NUMBER OF PASSES; MULTIPASS CROSS-COUNTERFLOW EXCHANGER $C_{min}/C_{max} = 1$; UNMIXED FLOW WITHIN PASES, ONE FLUID UNMIXED BETWEEN PASSES


To Stack
To Stream
At Least 8D
D
Access Door
To Afterburner
From Afterburner

A. Good Practice

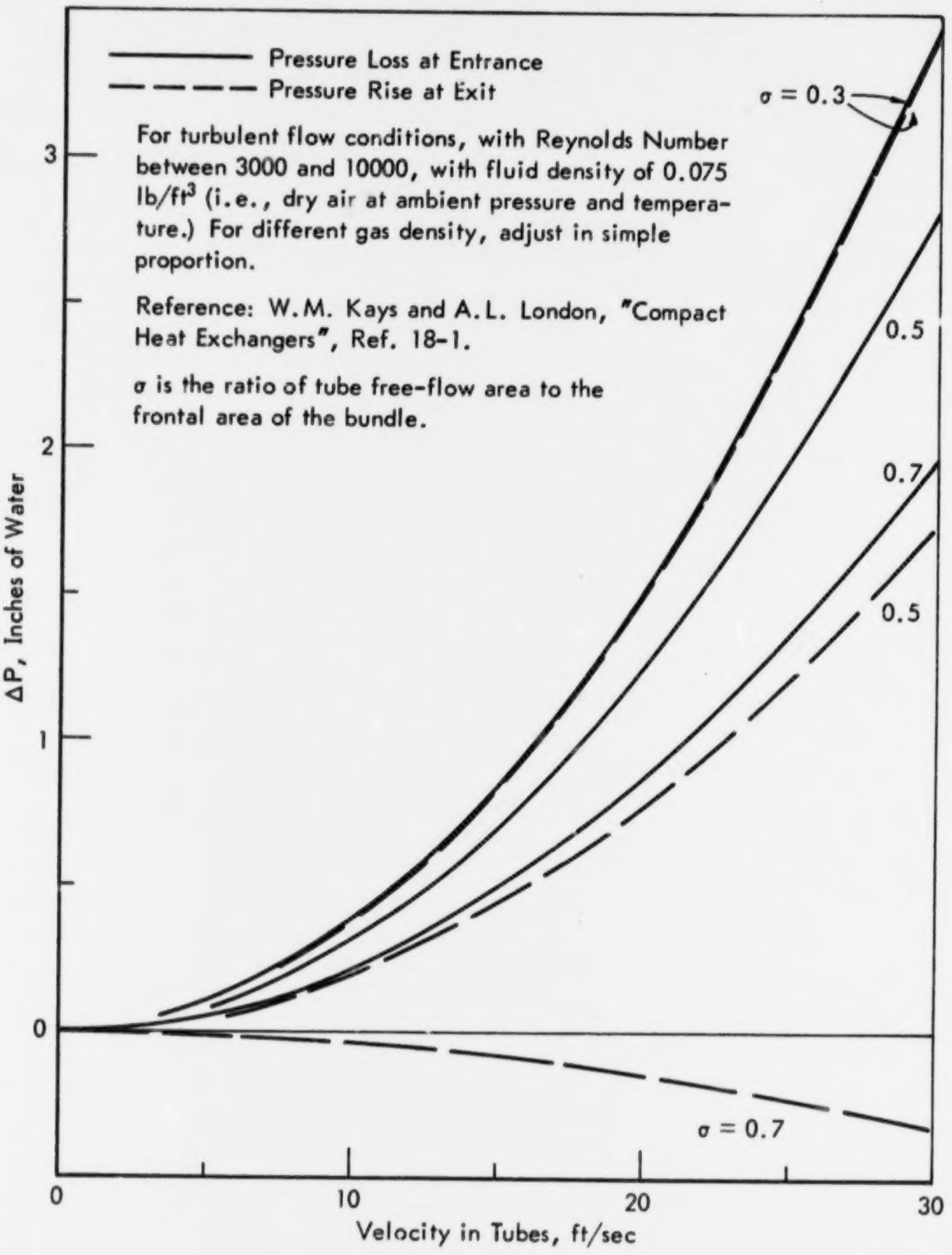

To Stack
Fume Stream
To Afterburner
Flue Gas

B. Good Practice



C. Poor Practice

Figure 18-8. HEAT EXCHANGER HEADERS



For turbulent flow conditions, with Reynolds Number between 3000 and 10000, with fluid density of 0.075 lb/ft^3 (i.e., dry air at ambient pressure and temperature.) For different gas density, adjust in simple proportion.

Reference: W.M. Kays and A.L. London, "Compact Heat Exchangers", Ref. 18-1.

σ is the ratio of tube free-flow area to the frontal area of the bundle.

Figure 18-9. ENTRANCE AND EXIT LOSSES FOR TURBULENT FLOW IN TUBES

For example, for air at atmospheric pressure with an approach velocity of 10 ft/sec and an area ratio, σ of 0.25 (defined below), the entrance pressure loss is .03-inch water.

The entrance pressure drop can be calculated for flow through the tubes by (Ref. 1, Chapter 5):

$$\Delta p = 0.003\ \rho V_1^2\ (1.5 - 0.4\ \sigma - \sigma^2) \qquad (18\text{-}12)$$

The exit pressure rise regained after flow through the tubes can be calculated from:

$$\Delta p = 0.003\ \rho V_2^2\ (2 + 1.9\sigma - 1.6\sigma^2) \qquad (18\text{-}13)$$

Here Δp is the pressure drop in inches of water; V_1 and V_2 are the approach and exit velocities in the header, ft/sec; ρ is the gas density (an average value through the exchanger can be used), lb/ft^3; and σ is the area contraction ratio between the header cross section and the tube openings

$$\sigma_{thru} = \frac{\pi}{4}\,\frac{n d_i^{\,3}}{A} \qquad (18\text{-}14)$$

where n = number of tubes
d_i = inside diameter of tubes, in.
A = the area of the tube plate exposed to the header, in.^{2}

Figure 18-9 gives a plot of entrance and exit Δp for a range of approach velocities and σ.

18.3.3 Friction Through Exchanger Tubes

This pressure drop can be calculated as follows:

$$\Delta p = 0.0011\ \frac{L}{d_i}\ \rho V^2 avg \qquad (18\text{-}15)$$

where Δp = pressure drop from tube wall friction, **in.** water.
L = the length of the tubes, in.
d_i = the inside diameter of tubes, in.
ρ = average gas density between inlet and outlet exchanger ft/lb
V_{avg} = average gas velocity inside the tubes, based on mean of entrance and exit densities, ft/sec.

(The equation is based on a Fanning friction factor of .008, which is in turn based on the assumption of Reynolds Number between 3,000 and 10,000, and on smooth tube walls; the friction factor becomes smaller at higher Reynolds Numbers, so Δp calculated is conservative. A more accurate calculation will account for variations in tube diameter and physical properties of the stream; as in Ref. 18-2.)

18.3.4 Pressure Drop for Flow Across Tubes

For flow across the tubes, entrance and exit effects are not accounted for separately. The entire flow channel through the exchanger is treated as a series of abrupt contractions and expansions.

For smooth round tubes with in-line arrangement:

$$\Delta p = 0.0010\ N\rho V^2_{max} \qquad (18\text{-}16)$$

If the tubes are staggered in a triangular pattern:

$$\Delta p = 0.0014\ N\rho V^2_{max} \qquad (18\text{-}17)$$

Here Δp is the pressure drop through the tube bank, inches of water, N is the number of tube rows deep through which the gases must pass, ρ is the average gas density in passing through the exchanger pass, and V_{max} is the velocity based on the smallest area between the tubes in the flow direction. Figure 18-10 illustrates the basis for determining this velocity. (The equations are based on equations given by McAdams Heat Transmission, **Ref.** 18-2, McGraw-Hill, for tube spacing of two diameters and Reynolds Number of 10,000. A more accurate calculation will allow for the effect of different tube layouts, and variations in fluid properties. See Ref. 18-4.)

18.3.5 Balancing Pressure Drop Against Exchanger Size

In selecting an exchanger, one often starts by specifying a desired effectiveness. Usually there is also a limited total pressure drop available for the complete system, which is determined by the flow resistance of afterburner, ductwork, heat recovery system, stack, and perhaps the oven or process unit as well, if all are driven by a single blower. The pressure drop assigned to the heat recovery unit, assumed here to be a cross flow heat exchanger, is thus a rather abritrary matter. It is a matter of economic balance to trade off size and cost of the exchanger that will achieve the desired effectiveness against the pressure drop that can be afforded. The key variable is the velocity used in the heat exchanger passes. The higher the velocity, the larger the value of the heat transfer coefficient, the smaller the passage flow areas, and the smaller the exchanger; the reduced exchanger size must be paid for by increased blower power. The pressure drop is very sensitive to the velocity, since it varies as the square of the velocity.

The process of design is a trail-and-error one, starting from a range of velocities, from which are calculated the exchanger sizes, effectiveness ratios, and pressure drops, so that one can establish which design meets requirements at the lowest cost. Ideally one could go to great lengths adjusting the many design variables to achieve a balanced and optimum design. Practically, the design is limited by the ready availability of standard designs, based on standard tubing sizes, tube layouts, frame sizes, etc., and a familiarity or predilection a given manufacturer may have for his own special design of flattened tubes, rectangular tubes with corrugations, etc., which are his stock in trade. Further, the manufacturer accumulates his own data and operating experience, which simplify his application.

Figure 18-10 follows

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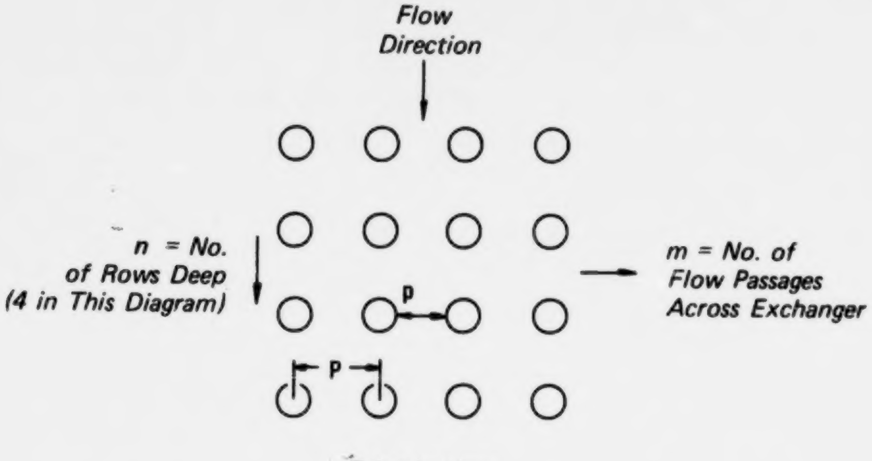
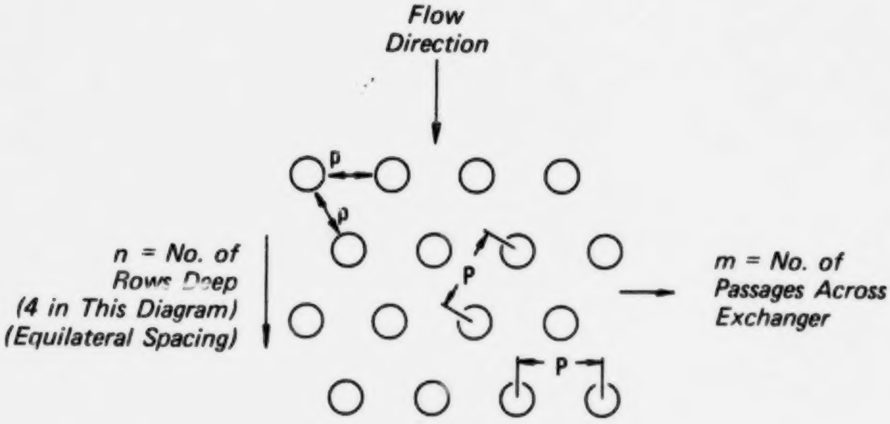

Flow Direction
n = No. of Rows Deep (4 in This Diagram)
P
P
m = No. of Flow Passages Across Exchanger

Tubes In-Line


Flow Direction
p
p
P
n = No. of Rows Deep (4 in This Diagram) (Equilateral Spacing)
P
m = No. of Passages Across Exchanger

Tubes Staggered

$$V_{max} = 0.2 \frac{Q}{mwL} = \frac{V_{avg}}{\sigma} = 0.2 \frac{Q}{mPL}$$

P = Tube to Tube Centerline Distance, in.

V = ft/sec

Q = ft^3/min at Gas Pressure and Temperature

m, n = Numbers of Tubes in Row

L = Tube Length, ft

p = Minimum Tube Clearance, ft

σ = (See Figure 18-11)

Figure 18-10. TUBE ARRANGEMENTS FOR CROSS FLOW

S-14121
67784

18.3.6 Other Miscellaneous Equations

Average Gas Density,

$$\rho_{avg} = \frac{M_{avg}}{10.72} \; \frac{(p + 14.7)}{(T_{avg} + 460)} \qquad (18\text{-}18)$$

Average Mole Weight

$$M_{avg} = \frac{1}{\frac{w_1}{M_1} + \frac{w_2}{M_2} \; . \; . \; .} \qquad (13\text{-}19)$$

$$\text{or } M_{avg} = v_1 \, M_1 + v_2 \, M_2 \; . \; . \; . \qquad (18\text{-}20)$$

where w_1, w_2 . . . = weight fraction (percent by weight/100) of components
1,2, . . . whose molecules weights are M_1, M_2 . . .
v_1, v_2 . . . = volume fraction (percent by volume/100 of components
1, 2 . . ., etc.

p = pressure of gas, lb_f/in^2, gauge

T_{avg} = average temperature of gas, °F.

ρ_{avg} = average density, lb_m/ft^3

18.4 Sizing an Exchanger - Procedure

The assumptions used in the following section are to be taken as suggestions, not prescriptions. It is suggested that Chapter 6 be reviewed before attempting to size an exchanger, since the general design considerations discussed there may be of overriding importance in choosing design assumptions.

Despite the direct relationship between E, R, and N_{tu} (given here in graphical form) the design problem is a trail-and-error one, since a number of constraints must be applied (available pressure drop, practical tube layout, etc.) which cannot be included in a direct relationship. Steps for sizing to meet heat transfer requirements are:

1) Select a desired effectiveness ratio.

2) Determine inlet, outlet, and average temperatures for both fume and flue gas streams (Equation 18-8).

3) Find inlet, outlet, and average gas densities for both streams (Equations 18-18, 18-19, or 18-20).

4) Select tube size, spacing, and arrangement.

Suggested Assumptions (Based on Typical Installations)

a) Fume stream passes through tubes
b) Flue gas passes across tubes
c) Tube diameter 1" OD, 0.868" ID stainless steel
d) Tubes staggered, on 60° centers, spaced 1-1/2" apart.

5) Using Figures 18-10 and 18-11, determine the free flow area ratios, σ_{cross} and σ_{thru}.

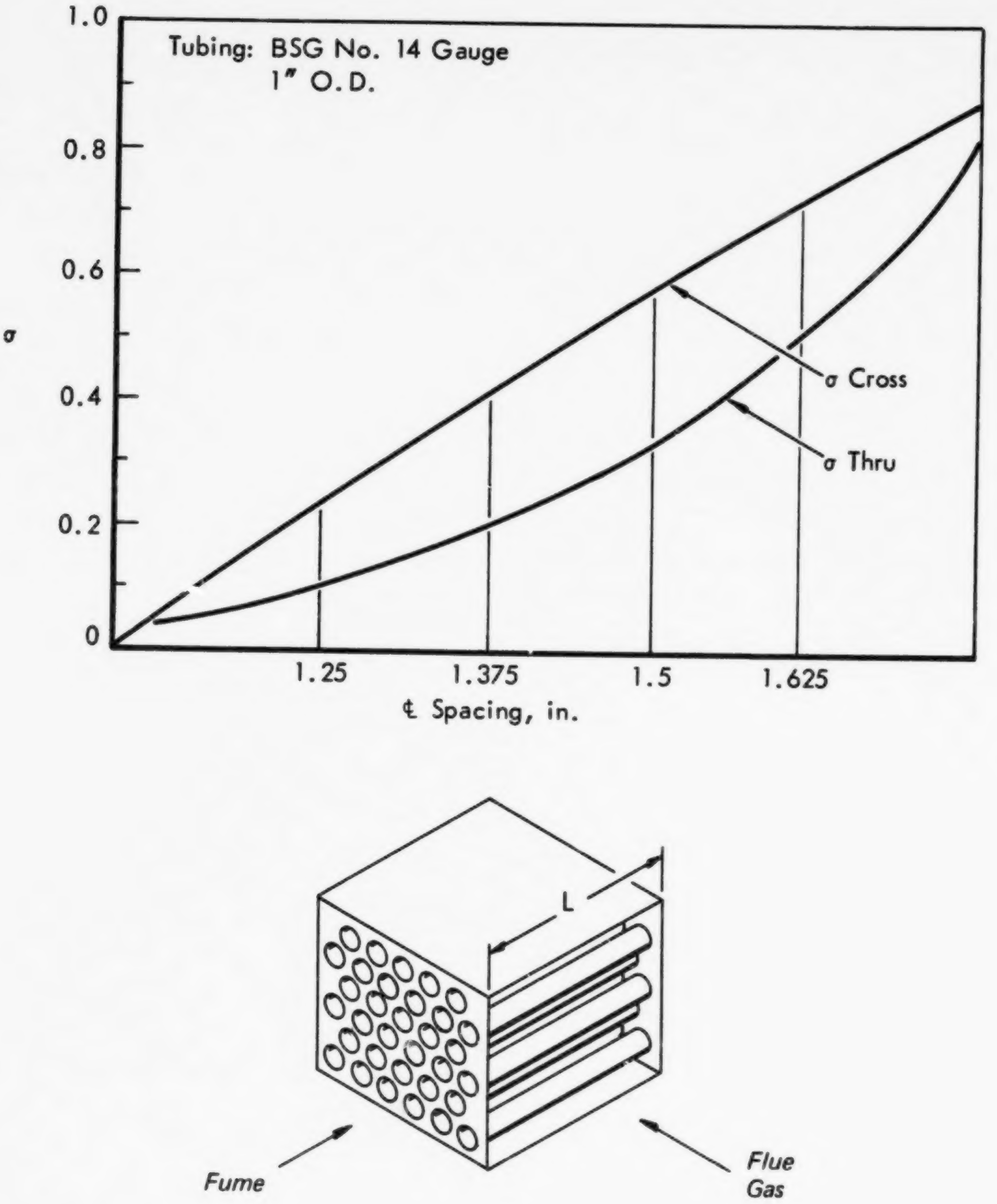
6) Assume a design velocity, $(V_{max})_{inside}$, for flow through the tubes. The pressure drop is a strong function of this velocity; it is suggested that two or three values be assumed, and that corresponding calculations be brought along. Suggested values: 40, 50, 60 ft/sec.

7) Based on this velocity (or velocities), using Figure 18-2, determine the inside heat transfer film coefficient, h_{fi}. Note that Figure 18-2 gives only an approximate value of h_{fi} based on an assumed tube diameter and gas composition; for more accurate values of h_{fi}, standard correlations based on Prandtl and Reynolds Number should be used, as given in Reference 18-2.

8) Assume fouling factor (see Figure 18-5). Suggested allowances are given in Table 18-2 for different services.

9) Determine wall conductance (Equation 18-6).

Figure 18-11 follows



$$\sigma = \frac{\text{Free Flow Area (Thru or Across Tubes)}}{\text{Frontal Area (Approaching Exchanger)}}$$

Figure 18-11. FREE FLOW AREA FOR FLOW ACROSS A TUBE BUNDLE

10) Assume a design velocity, $(V_{max})_{outside}$, for flow across the tubes. In order to make the pressure drop roughly equal in both through- and cross-flow passes, it is suggested that the velocity be taken:

$$(V_{max})_{outside} = C \sqrt{\frac{(T_{avg})_{outside}}{(T_{avg})_{inside}}} \times (V_{max})_{inside}$$

where C = 1.1 for in-line tubes
= 0.9 for staggered tubes

The term $\sqrt{\frac{(T_{avg})_{outside}}{(T_{avg})_{inside}}}$ allows for the different gas densities in the passes; it is based on the absolute temperature, °Rankine (°Fahrenheit + 460). For a cross-flow exchanger having an effectiveness of 45% and a flue gas temperature of 1400°F, the term has a value of about 1.4.

11) Determine the outside heat transfer film coefficient for flow across the tubes using Figure 18-3, ρ_{avg} from step 3, and $(V_{max})_{outside}$ from step 10. As for step 7, this value of h_{fo} is approximate, and more accurate values can be found by referring to the generalized correlations of Kays and London (Ref. 18-1), McAdams (Ref. 18-2), or Bell Ref. 18-4). In particular, with some tube arrangements, the value of h_{fo} will vary with the number of tube rows deep (n in Figure 18-11).

12) Determine overall heat transfer coefficient (Equation 18-7).

13) Find average specific heat (Figure 18-1, taken at average fume temperature through exchanger).

14) Determine mass flow rate of fume through tubes (lb/hr).

15) Determine mass flow rate of flue gas = fume + fuel + combustion air (if any is added), in lb/hr.

16) Determine N_{tu} (Figure 18-6) using selected effectiveness from step 1 above.

17) Find $R = C_{fume}/C_{flue\ gas}$ (Equation 18-11). Flow through tubes is "unmixed"; flow across tubes is taken as "mixed".

18) Calculate the required inside tube area (Equation 18-9 and Steps 16, 12, and 13).

19) Calculate the number of tubes required, using tube inside diameter, V_{max} from step 5, mass flow rate, and average gas density (Step 3).

$$N = \frac{(144/60)w}{(\pi/4)d_i^2(V_{max})_{inside}} = 3.06 \frac{w}{(d_i)^2(V_{max})_{inside}}$$

20) Calculate the required length of tubes, L, using Equation 18-10. Adjust if necessary to use standard tube length.

21) Determine the required area for cross flow over the tube bank to carry the required flue gas, using $(V_{max})_{outside}$ from step 10, ρ_{avg} from Step 3, and the mass flowrate of the flue gas. This is the net free flow area between the tubes.

$$A_{net\ cross\ flow} = \frac{w}{60\ \rho_{avg}(V_{max})_{outside}}$$

22) Calculate the number of tube rows across the exchanger (perpendicular to the flow direction), m, using Figure 18-10 (which defines the free space width between the tubes, ρ), the area from Step 21, the length L from Step 20.

$$m = \frac{A\ cross\ flow}{\rho\ L} - 1$$

23) Determine the number of tubes deep, n, in the direction of flow, using the total number of tubes from Step 19.

$$n = \frac{N}{m}$$

The number of tubes, N, their length, L, and the number of tube rows across and thru, n and m, should be adjusted to achieve integer values for m and n.

24) The corresponding header frontal areas, A header, of the exchanger may be found using the values of σ from Step 5. The proportions of the headers are fixed by the number of tube rows across, m, and deep, n, and the tube length, L.

18.5 Procedure for Checking Pressure Drop

The steps above have sized the exchanger--tube diameter, tube spacing, number of tube rows deep and wide, and tube length. It remains, now, to determine pressure drops. This is done as follows:

1) Determine gas density at entrance and exit conditions in the headers (Equation 18-18).

2) Determine approach and leaving velocities $V_{avg} = Q/A_{header}$.

3) Determine tube entrance loss and exit gain (Equation 18-12, 18-13).

4) Using average velocities and densities through both passes, determine friction pressure drop;

a) Through tubes, (Equation 18-15).

b) Across tubes, (Equations 18-16, 18-17).

Note that the values for f given in sections 18.3.3 and 18.3.4 are approximate, and reference 18-4 should be used for more accurate calculations.

5) Sum the pressure drops (with due regard to pressure rise at tube exit). Pressure drop through the two passes should be roughly equal. If not, re-adjustment of the number of tube rows--width and depth--may be needed, keeping the total number of tubes and area constant.

6) The total pressure drop should be compatible with the combined system and the blower supply pressure. If not satisfactory, the design should be repeated with higher or lower velocity assumptions (Setps 6 and 10 in the heat transfer design).

18.6 Evaluating the Performance of an Existing Exchanger

This is a much more direct problem than that of sizing the exchanger, which involves successive trial calculations to arrive at an acceptable design. It is assumed that the important design and operating variables are fixed:

1) The dimensions of the exchanger and its ducting,

2) The entering fume stream and flue gas temperatures,

3) The corresponding flow rates and compositions.

One may proceed to calculate the effectiveness, the exit gas temperatures, and the pressure drops. Steps are not outlined in detail; the first step is to assume approximate exit gas temperatures and determine average gas densities and specific heats (subsequent recalculation will not be necessary in many cases, since the changes in density and specific heat are not large). Then average velocities and heat transfer coefficients can be found. The number of transfer units and the capacity rate ratio may then be calculated, and the effectiveness determined. The effectiveness allows recheck of the assumed temperatures. Calculation of pressure drops is straightforward.

Chapter 19. ANNOTATED LIST OF EQUIPMENT SUPPLIERS

The following listing of afterburner manufacturers and associated equipment suppliers is only partly complete. Annotations cannot give a complete resume or a fair presentation of a given manufacturer's capabilities. The field is a rapidly changing one, and equipment supplied, and the manufacturers engaged in business change continually. However, many of the companies listed have been designing and manufacturing afterburners and other incineration equipment for periods of 5 - 20 years or more. The listing is included primarily to aid a potential afterburner user in starting to look for the equipment he wants.

The authors of this report were unable to visit and discuss details with all of the equipment suppliers. Also, some were more willing than others to discuss details of design and design procedures. As a result more space has been spent describing the products of some suppliers than of others. This should not be considered as a reflection on the relative merits of one supplier over another, or of their respective products.

Afterburners - Thermal

AER Corporation, Ramsey, New Jersey, 07446. This company makes custom designs based on the Maxon "Combustifume" burner and box construction lined with fibre block insulation. They will supply tubular crossflow heat exchangers. Applications have been mostly in paper and fabric industries.

Air Preheater Company, Wellsville, New York, 14895. They manufacture pre-engineered units in the 2000 to 30,000 scfm range as well as custom designed units. These are mainly straight-through horizontal afterburners, light in weight (insulated with soft block and fiberfrax); they are available fired either with gas (using the Maxon "Combustifume" burner) or with oil (No. 2, No. 4, or No. 6 fuel oil, using a proprietary Air Preheater mixing region together with a North American dual-fuel burner). Heat exchange units are emphasized, using the proprietary "Cor-Pak" exchanger (corrugated tube surfaces), with recuperative heat recovery of 45 to 50% and 65 to 70% available as secondary recovery using a rotary regenerative exchanger. Air Preheater has carried out applications in odor elimination in rendering plants, fume and odor elimination in rubber curing, fume and odor problems in fiber glass impregnation, particle board curing, and a number of more conventional solvent evaporation processes. They maintain test facilities including a 2000 scfm prototype at their plant, and have available a 100 scfm portable unit for on-site testing.

Alkar Engineering Corporation, 105 Spring St., Lodi, Wisconsin 53555. Their basic design is a horizontal venturi-shaped chamber, axially fired with a Maxon "Ovenpack" burner, and with radial fume entrance at the burner end of the chamber. The chamber is lined with medium, low density castable refractory. They offer a recuperative heat exchanger in connection with this design, but only one has been installed. They have extensive experience with the special application problems of afterburners for smokehouses (e.g., fouling of fans and ducts with fume condensates), and a large fraction of their business is supplying smokehouse equipment to meat processors.

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Bayco Industries of California, 470 Hester St., San Leandro, California. Bayco builds small package units and custom designed units of larger size. Their package units are rectangular boxes with U-bend and stack, lined with castable refractory, and a small straight stack unit. They use a variety of burners. Small units are used for smokehouses, electrical insulation burnoff, sewage odor control. Custom units are used for burning oil vapor off filter recovery units, and for metallurgical furnaces. They have no experience with heat recovery.

Bigelow-Liptak Corp., Northwestern Highway and 10-1/2 Mile Road, Southfield (Detroit), Michigan 48075. Their equipment, custom-designed for each application, characteristically involves heavy-duty, brick-lined incineration units. They are aiming primarily for the large-scale and special conditions applications rather than pre-engineered package units. They have designed units as large as 150 M scfm, utilizing waste heat boilers (rather than recuperative heat exchangers) for heat recovery. They have their own design for distributed raw gas burners. Bigelow-Liptak have experience supplying equipment for incineration of liquid, solid, and gaseous wastes. This experience includes such difficult materials as streams containing halogens, other inorganics, and particulates. For such applications, they have designed systems including pre- and post-treatment of the stream as well as incineration.

Blaw-Knox Company, 1575 Fillmore Ave., Buffalo, New York 14211. Blaw-Knox offer an afterburner designed and manufactured in conjunction with Thermal Research and Engineering for moderately-large, continuous coffee roasting equipment, and one of their own design for use on smaller, batch equipment. Neither unit involves heat exchange, both are relatively short stack incinerators. Their experience is in manufacturing afterburners in connection with food processing machinery, primarily for coffee roasters.

Clean-Air-Rator Co., 1940 Linwood Blvd., Oklahoma City, Oklahoma 73106. This company builds stack-type units with metal combustion chamber lining, bottom-fired, with package burners using gas fuel. They have been used in food processing and refuse incineration applications.

Coats Warner Co., 15130 Ventura Blvd., Sherman Oaks, California. They build gas-fired afterburners, custom designed, primarily in conjunction with their paint handling equipment, 750 to 1500 scfm, to LACAPCD standards. Recycle flue gas to ovens has been used for heat recovery.

Combustion Equipment Associates, Todd Division, 120 Park Ave., New York City 10017. Their thermal afterburner units are custom designed for relatively large applications, and include combination gas/oil fired burners. Their technical expertise is focussed on design of an effective, light-weight combustion chamber. Heat recovery is not stressed and standard equipment of other manufacturers is used for this purpose. They have limited experience with fume incineration, but broad experience in combustion and thermal incineration of liquid and solid wastes.

Despatch Oven Company, P.O. Box 1320, Minneapolis, Minnesota 55440. Their basic design is a horizontal-folded chamber with a Maxon "Combustifume" burner (gas firing) at the inlet. Units are insulated with low density castable refractory and hence are relatively light weight - they are also relatively low cost. Despatch maintains laboratory and proto-type test facilities at their plant, including a 3000 scfm unit, to develop designs and investigate application problems. Their business is primarily the manufacture of industrial ovens and they have entered the fume incineration business in order to treat effluents from their ovens. They offer a set of nine sizes of a standarized design with or without heat exchanger ranging in capacity from 400 to 14,000 scfm. Over 90% of their afterburners are sold in conjunction with new oven installations, mostly for paint spraying or baking applications.

FECO, 5855 Grant Ave., Cleveland, Ohio 44105. Afterburners available from FECO include conventional (brick or castable lined) thermal afterburners, with or without tubular cross-flow heat exchangers. Units are custom designed with either rectangular or circular cross sections with brick checkerwork baffles to promote mixing, and are either gas or oil fired. They also make economy, light-weight thermal units, using silica fiber insulation, and supplied with or without rotary wheel regenerative heat exchangers. These units are of rectangular cross section with a folded (U-bend) flow path. They are designed for 0.5 to 0.9 second residence time and flow velocities of 25 to 30 ft/sec at 1500°F. They are designed for natural gas firing only, using a distributed raw gas burner. FECO also makes catalytic units (see section below). FECO's principal business is the manufacture of ovens for metal decorating and foundry core baking. They have added afterburners to their product line primarily in order to supply complete installations to purchasers of their ovens.

Frimberger Corp., 15301 12 Mile Rd, Rosevillem Michigan 48066 (formerly Mahon Industrial Div.). This company builds fume incinerators, dust collectors, mist eliminators, and precipitators.

Gas Processors, 404 No. Berry St., Brea, California 92621. Standard units, 250 - 10,000 scfm are supplied with gas burner and have a horizontal cylindrical holding chamber, lined with castable refractory. They will supply units with steam boiler for waste heat recovery. Their experience has been chiefly on paint baking ovens and spray systems, and on odor elimination.

Granco Equipment, Inc., 1958 Wilson Ave. S.W., Grand Rapids, Michigan 49504. This company uses a metal U-tube combustion chamber; incoming fume stream surrounds the U-tube and exchanges heat; metal fins increase heat transfer area. Design uses "Flame Grid" burner. Their first installation is undergoing test.

Hayes Albion, 8550 W. Michigan Ave., Parma, Michigan 49269. This design is a vertical metal stack unit with distributed gas burner using outside air and a second manifold for introducing fume. It has been applied to pulp mill digester vapors for odor control.

Hirt Combustion Engineers, 931 So. Maple Ave., Montebello, California 90640. Hirt provides both standarized and custom designs. The combustion chamber uses a castable refractory lining inside a cylindrical metal shell. Blowers operate on exit gases to minimize the possibility of fouling; air dilution is used at the blower entry to reduce the temperature. Hirt's gas burner uses a multiple array of ceramic cup-shaped burner heads, each with its own gas jet, and operated at near-stoichiometric conditions. It gives a turndown of about 10:1, and groups are sectionalized if more turndown is required. Provided it is not a fouling stream, or deficient in oxygen, the fume air is used in part for combustion; the remainder is introduced after the burner assembly for dilution and cooling to the required holding section temperature. A throat ring and ring-and-donut baffles are used to insure good mixing in the holding section. Hirt's oil burner (their own design) is steam or air atomized. Because of its lower turndown, several burners may be used. Hirt use standard designs, but give an engineering review of the process. They have supplied thermal afterburners to a wide variety of applications, many of which have included heat recovery, using either tubular exchangers or rotary recuperators.

Koch Engineering Co., Inc., 41 E. 42nd St., New York, New York 10017. Koch supply custom-designed afterburners, based on commerically available burners, using refractory-lined combustion chambers.

T. Melsheimer Co., Inc., 22610 So. Western Ave., Torrance, California 90501. This company has thermal incinerators (8000 and 15,000 scfm) for sewage plant odor control. Vertical fire-tube heater units are used for exchange-preheating the fume stream.

Michigan Oven Co., P.O. Box 336, Romulus, Michigan 48174. Afterburners are built for use in conjunction with charring ovens (wire insulation burnoff).

Mid-South Manufacturing Co., 1221 Bank of the Southwest Bldg., Houston, Texas 77002. They use a small concentric-tube metal chamber with internal baffling, 150 - 1300 scfm, temperature 1100°F; stack gas exchanges heat with fume stream. Units have been used for sewage lift station odor control.

B. Offen and Company, 29 East Madison St., Chicago, Illinois 60602. They build custom units using the gas-fired Maxon "Combustifume" burner in rectangular or cylindrical chambers (depending on size and where it is to be located). They have used tubular exchangers for heat recovery. Offen's experience have been in supplying equipment primarily to the graphic arts industry for use with printing presses.

Pacific Coast Incinerators, 710 Delaware St., Berkeley, California. They use a vertical cylinder with castable refractory lining, and a package gas burner. Units have been supplied for restaurants, paint-bake ovens.

Proctor and Schwartz, Incorporated, Pollution Control Equipment Division, 7th St. and Tabor Rd., Philadelphia, Pennsylvania 19120. They have developed a novel regenerative heat exchange technique coupled with thermal fume incineration, which is the basis of their "TRAPS" system. They have one such unit in operation, and have designed the system for simple operation and modular construction. The first unit of this new design is now being installed. The system allows a high degree of heat economy, and has been developed through use of a pilot scale unit. This novel approach does require a cyclically operated valve arrangement, and uses ceramic structural members operating at high temperatures.

Ross Engineering Division, Midland-Ross, P.O.Box 147, New Brunswick, New Jersey 08903. Ross Engineering design and supply medium to large thermal afterburners with emphasis on heat recovery through both preheating of the contaminated air stream and heating a secondary fluid system. They have experience with normal solvent systems and maintain a test facility at New Brunswick. Their normal design is a downfired rectangular unit using a Maxon "Combustifume" burner fired under a bridge wall and through a heat exchanger. They have also designed end-fired cylindrical units using both gas and oil fired burners. Their units are designed for low overall pressure drop, and use a metal structure with insulating masonry lining. Applications have been for resin and paint curing and drying ovens.

J. B. Seage Co., 2419 E. 56th St., Los Angeles, California 90058. Custom-designed units use a vertical stack, are refractory lined; fume enters at the bottom, gas burner (package unit) enters at side. Units are in service in curing ovens, crematories, metallurgical furnaces, burnoff ovens.

Services and Equipment Co., 4401 S. Western Blvd., Chicago, Illinois 60609. They build small, inexpensive stack-type units, with side entry of gas flame, short residence time. Have been used for coffee roasters, restaurant boiler vents, wire burnoff ovens.

Smith Engineering Co., 1903 Doreen Ave., South El Monte, California 91733. Their designs include horizontal, vertical, induced and forced-draft types; units are custom designed. Induced draft designs are preferred. Their designs are generally brick lined, use a distributed raw gas burner of Smith Engineering design, and stainless steel construction in burner and heat exchanger (when used). Larger units use masonry construction, smaller units use castable refractory. Experience includes paint baking ovens, coffee roasters, air sterilizers, resin curing ovens.

Spencer Boiler and Engineering Company, 12106 So. Center St., Southgate, California 90280. Units are of standard design, but are supplied as custom applications. They use a multijet gas burner of their own design and a cylindrical holding chamber, lined with castable refractory. They have used waste heat boilers, exchangers, water heaters, and recycle gas streams for heat recovery, and have applied units to smokehouses, paint and solvent areas, and food dehydrating and roasting.

Surface Combustion Division, Midland-Ross Corporation, P.O. Box 907, Toledo, Ohio 43601. Their afterburners are standardized, pre-engineered units which cover low and middle-range capacities (100 to 10,000 scfm), and among the four types cover rich fume streams as well as more common lean streams. All units are gas-fired. Primary emphasis is on adequate design of the combustion chamber and low pressure drop through the units; no standard heat exchanger design is offered. Larger units, including heat exchange would be designed on a custom basis. The test facilities used in the research program, including a 3000 scfm prototype, are maintained to study unusual applications problems. They have carried out an extensive research program in developing designs for the four types of afterburners which they offer. Published results of this program represent an important contribution to defining the fundamental requirements for thermal afterburner design.

Tailor and Company, Inc., Bettendorf, Iowa. This company have built chemical incinerators, but have no experience in fume afterburners. They use a unique design of cyclone burner.

Thermal Research and Engineering, Conshohocken, Pennsylvania 19428. Their units are custom designed, tend to be for medium to large applications; many feature a very high degree of recuperative heat exchange, and the use of a combination fuel oil/natural gas fired burner. Several different burner/combustion chamber configurations are used, depending upon the application. Burners are of Thermal Research design. They have a broad range of experience, and maintain a flexible test facility used in developing designs for new applications. Applications have included chemical and organic wastes, smelter off-gases, sulfur plant tail gases, chlorinated wastes, tars. They have tended to concentrate on unusual wastes where experience and engineering skills count for more than equipment cost.

UOP Air Correction Division, Tokoneke Rd., Darien, Connecticut 06820. UOP build a range of pollution control equipment (e.g., vent scrubbers, cyclones, precipitators, adsorption and absorption equipment) as well as both thermal and catalytic afterburners. Both thermal and catalytic unit designs have employed lightweight, all-metal construction, using stainless steels without refractory lining. Current designs use refractory board linings. They have pre-engineered units in sizes to 6,500 scfm. Larger units may incorporate heat exchangers. They normally use a distributed raw gas burner with a perforated conical baffle for fume inlet, of UOP design. UOP experiences have included industrial chemicals and oil refining applications, paint and varnish cookers, sewage vent treating.

John Zink Company, 4401 So. Peoria St., Tulsa, Oklahoma 74105. Zink Co. design and build custom engineered installations, called "Thermal Oxidizers". They use varied arrangements depending upon the job. One arrangement is a squat vertical cylinder, joined by a conical section to a vertical stack. Opposed burners fire radially into the cylindrical part, and are without internal baffles. On waste streams not previously encountered, they run a pilot test. Halogenated streams are sometimes given as much as 3 seconds residence time. Zink makes their own burners, which can be supplied for gas or liquid fuels, or both, and follow practices on process heaters. They commonly use heavy masonry construction. Applications have largely been in industrial chemicals and refining plants. Units tend to be of large size, may be used as fume oxidizers or as gaseous or liquid-fuel incinerators; they may be part of a system that includes scrubbers, knockouts, and precipitators when waste streams are halogenated (or otherwise corrosive or objectionable). They have not had experience with installations using heat exchangers.

Vari-Systems, Environmental Systems Division, 1295 W. 78th St., Cleveland, Ohio 44102. Standard models, oil or gas fired, use fume or outside air for combustion. Refractory-lined horizontal cylindrical chamber are end fired or have double-wall construction. Custom units will be designed for large sizes or special duties.

Afterburners - Catalytic

Englehard Minerals and Chemicals, Gas Equipment Section, 205 Grant Ave., East Newark, New Jersey 07029. They will provide catalytic afterburner equipment for oxidation of waste fume streams. They have considerable parallel experience in catalytic units for reduction of nitrogen oxides, but limited experience with oxidation units. Their basic design is a downfired preheat burner, with heated gases flowing under a bridge wall and up through the catalyst matrix. Englehard uses noble metal catalysts supported on a honeycomb ceramic grid "Thermocomb" supplied by American Lava.

E. I. du Pont de Nemours and Co., Industrial and Biochemicals Department, Wilmington, Delaware 19898. Du Pont supplies the honeycomb ceramic matrix material, "Torvex", to other catalytic afterburner manufacturers, and supplies noble metal catalysts based on this material both to users and manufacturers of catalytic afterburners. Their development group has done extensive work on catalyst development and characterization, and is available as consultants in design and application of afterburners.

FECO, 5855 Grant Ave., Cleveland, Ohio 44105. Catalytic units are custom designed, using ceramic honeycomb-based catalysts supplied by du Pont or Matthey-Bishop. Normal designs use a rugged, double-wall construction with stainless steel inner wall and heavy gauge carbon steel outer wall with ceramic fiber block or wool between for the enclosures and distributed raw-gas burner for preheat. As with their thermal units, most applications have been in supplying air pollution control systems to FECO's oven customers.

Matthey-Bishop, Malvern, Pennsylvania 19355. Their catalysts are noble metals on du Pont "Torvex" honeycomb supports. A modular approach to mounting the catalyst element allows effective utilization in afterburner units with either rectangular or circular cross-sections. They also have design concepts and capability for design and construction of thermal afterburners, but have not yet supplied any. They have supplied catalytic units for reduction of NO_x in nitric acid plant tail gas. They have designs for catalytic systems for hydrocarbon and odor abatement, and have supplied replacement catalysts for existing catalytic (oxidation) afterburners. To date, their emphasis has been on relatively large installations in the wire-enamelling, metal decorating, printing and food processing industries, and for odor problems.

Oxycatalyst, Inc., E. Biddle St., West Chester, Pennsylvania 19380. Both catalytic and thermal afterburning equipment for mobile and stationary sources. Their early units were based on using "Oxycat" (brick) catalyst elements. These were arrays of ceramic rods with noble metal deposited on their surfaces, and fumes were passed in cross-flow over these rods in the array. "Oxycats" are still manufactured for replacement and a few new applications. Newer Oxycatalyst units use a noble metal catalyst on a du Pont "Torvex" honeycomb base. They have pre-engineered afterburner designs for flow rates between 2.5 and 12 M scfm as well as custom designed units for larger capacities. These units generally use distributed raw-gas burner to supply preheat and have a castable-refractory-lined enclosure of rectangular cross-section for the preheat burner, mixing and catalyst sections. They include heat exchangers when required. Their design configurations use single unit construction and are shipped as pre-packaged units. They will furnish units intended for use as catalytic afterburners, but which have the capacity (sufficient residence time, velocity and temperature resistance) of being converted to thermal afterburners. Such convertible units have a higher initial cost than straight catalytic units, but allow for the possibility of catalytic operation being inadequate either because of unforeseen problems in the application or because of changes in the nature of the user's fume stream composition or conversion requirements. Oxycatalyst has experience in a wide range of applications of catalytic afterburners.

UOP, Air Correction Division, Tokoneke Rd., Darien, Connecticut 06820. The catalyst used in catalytic units is the basic Suter-Ruff type with spongy noble metal deposited on nichrome ribbon, which is crimped and assembled in the unit as a mat in a module similar to a furnace air filter. Catalytic units are now generally of an inline arrangement of preheat burner and catalyst element rather than the folded configuration which has been much described in the literature. Enclosures are of light-gauge, double metal wall construction, with aluminized steel normally used as the inner liner. UOP's distributed conical burner is normally used for preheat. UOP has extensive experience in a number of afterburner applications, and has available rate constants for both thermal and catalytic oxidation of a number of typical fume constituents. These are used in proprietary computer programs which calculate temperature and time requirements for designs.

Burner Manufacturers

All of the major burner manufacturers make gas and oil burners which can be used in afterburners currently on the market. The four manufacturers in this list have had extensive experience with fume incineration and all have developed distributed burners.

Eclipse Fuel Engineering Company, 1100 Buchanan St., Rockford, Illinois 61101. They supply gas and distillate oil burner packages. They have a distributed source "line type" gas burner with provision for using the fume for the burner oxygen supply if the fume is non-fouling.

Hydro Combustion Corp., 22805 So. Avalon Blvd., Carson, California 90744. Although they have supplied thermal fume incinerators, their principal specialty is burners--they make a distributed gas burner and a heavy oil burner.

Maxom Corporation, Muncie, Indiana 47302. Maxon manufactures gas burners and auxiliary combustion equipment (combustion air blowers, controls, etc.). Their raw gas, "Combustifume" burner is widely used in fume afterburners. The "Ovenpak" premix gas burner is a package unit with blower and controls. The "wide range" type provides for a turndown of 40 to 1 on firing rate, using a proportioning air/fuel controller. They have carried out extensive research and development work on the time-temperature requirements for afterburning fume streams containing a variety of contaminants. They maintain a development test facility at Muncie and are continuing work on specialized application problems.

North American Manufacturing Company, 4455 E. 71st St., Cleveland, Ohio 44105. They are manufacturers of burners and auxiliary combustion equipment for both gas and oil firing. Their "dual-fuel" burner line is supplied where oil firing or optional oil/gas firing is required. They make a distributed raw gas burner, termed "Flame Grid", developed specifically for fume afterburner application, which can make use of the fume as an oxygen source for combustion. They supply a "high excess air" burner which will tolerate several hundred percent of stoichiometric air (using internal bypassing).

Chapter 20. REVIEW OF TASKS

The purpose of this section is to briefly describe the methods used in carrying out the afterburner systems study. The study covered afterburner or fume incinerator technology for control of gaseous (or gasborne particulate) combustible emissions from stationary sources. Disposal of liquids or solids by combustion was not included. Two main classes of afterburner were considered, thermal (direct flame) and catalytic. Flares and afterburners which were part of a solids refuse disposal system were not included. The objectives of the system study were to 1) evaluate present status of technology, 2) evaluate existing afterburner systems, 3) determine current practices and problems in applications, 4) identify potential future applications, and 5) develop R and D plans to advance technology. Information was obtained from the published literature, equipment manufacturers, equipment users, air pollution control agencies, research institutes and universities. Visits were made to many of these sources of information in order to hold detailed technical discussions and observe equipment in operation. Questionnaires were sent to plants which were known to have afterburners in use or were likely to have them. The respondents provided information on general performance, costs and problems encountered with their units. Detailed performance data were rarely available.

20.1 Evaluation of Current Engineering Technology

The current status of theoretical and applied technology applicable to catalytic and flame afterburners was reviewed. The first source of information examined was the published literature. While the combustion literature is very large, only a small amount could be found on afterburners. The researchers at universities and research institutes have been concentrating on combustion phenomena of fuels rather than waste destruction. The theoretical principles are, however, the same and this literature is utilized in this report. Even for fuel combustion, the theoretical aspects are not advanced far enough to be able to use them in design calculations for equipment. The bulk of the published literature has been written by the equipment manufacturers and is often only descriptive material and occasionally contains empirical calculation procedures. There is a little literature written by the users of afterburners but again this is quite general and empirical in nature. Part of the difficulty in developing a technological base is the lack of good test data on performance. Many early applications of afterburners were for odor and smoke abatement where no quantitative performance is defined. The second source of information was the equipment manufacturers, a number of whom were visited by members of our staff. A visit typically was arranged after a letter contact outlining the objectives of our study and listing the information we were looking for. The actual visit was made by two staff members and usually lasted about one day. Reports of these discussions were then written subsequent to the visit. These visits enabled us to obtain a clear picture of the technology in use for equipment design. While principles were always freely discussed, in some cases specific parameter values used in design were considered proprietary and were not revealed to us. By and large, designs have evolved in an empirical way and mistakes are still made through lack of understanding. Nevertheless, we were able to learn the principles of the approach taken by the manufacturers. The proprietary material is available to the potential user since it will be brought to bear on his problem when he approaches the manufacturer for equipment selection and purchase. The technology is reported in various parts of this handbook. We have brought together the theoretical background in fuel combustion with the empirical methods used in afterburner design. This provides insight for many of the problems encountered and points the way towards improvements.

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20.2 Evaluation of Existing Afterburner Systems

A review of the features of existing afterburner systems is reported. The main sources of information here were the equipment manufacturers. Their brochures and published literature contains this material in a readily available form. This information was supplemented by our staff visits to the manufacturers. Additional information on specific features of existing systems was obtained from users (the user survey is described in the next section). The advantages and disadvantages of the various features are presented above.
Some guideline information on operating and capital costs was obtained and have been summarized. This economic data can be used by a prospective user for orientation, preliminary selection of equipment and an approximate capital and operating cost. It is envisioned that more accurate costs would be developed in the specific dealings a user would have with an equipment manufacturer. A listing by manufacturer of equipment types available is contained in Chapter 19.

20.3 Assess Present Practices and Problems

20.3.1 Survey of Users

A survey of users of afterburners was made by sending out questionnaires to plants which were known to have afterburners, or it was thought likely that they would have them. Follow-up visits were made in some cases, and telephone calls in others, to obtain further information.

Response to the questionnaires was disappointing. Out of 559 questionnaires sent out, completed forms were received from 103 plant locations covering 238 afterburners. Although some companies responded promptly, in many cases it was necessary to make repeated telephone calls and send out duplicate questionnaires before replies were received. In a number of cases, information promised some time ago has not been received.

Data from the replies have been tabulated in summary form. A computer format was used in order to facilitate manipulation and machine plotting of the information.

20.3.2 Contacts with APCD's

Local air pollution control districts were contacted as part of our survey to:

1) Identify present afterburner users, especially those encountering difficulty with their units.

2) Obtain test data on the effectiveness of afterburner installations.

3) Obtain pollutant inventory data and learn of present and projected regulations governing hydrocarbon and odorous emissions.

Regional offices of APCO were contacted to determine the most active local districts which we then contacted, mostly through telephone calls. In most areas of the country outside of California, there has been little concern with pollutants other than smoke and SO_2. Therefore, most of the local APCD's had little to contribute to our study.

Where the air pollution control district indicated that they had afterburners in operation, indicated that they had experienced some problems, and where we were planning a trip to the area to visit an afterburner manufacturer, we conducted discussions with district personnel. Especially useful discussions were held with districts in the San Francisco Bay Area, St. Louis, and New Jersey. Eight other local and state agencies were visited. The Los Angeles County APCD, which has the most extensive test data and experience with afterburners discussed philosophy of afterburner design,testing, and regulations but names of users and test data were considered proprietary. This concern over release of user names was widespread. Test data is almost non-existent, but we were supplied much of that obtained by the Bay Area APCD after user names had been removed. The St. Louis district took us to visit several afterburner installations, but felt that all local user contact should be handled through them.

20.3.3 Questionnaire Format

The objective of the questionnaire was to obtain information on the various afterburners being used in industry so that performance and cost could be evaluated. This presented a problem as to what questions should be included and how long the questionnaire should be.

A draft questionnaire was submitted for approval in early October, 1970. This was later amended, and the final version was given Bureau of Budget approval on December 29, 1970. A copy of the final version, together with the cover letter is attached. In final form the questions cover six single side pages, and the length may have contributed to the poor response.

20.3.4 Selection of Users

Names of users were first of all obtained from lists supplied by afterburner manufacturers and by Air Pollution Control Districts. Most of these addresses were incomplete (and sometimes wrong), so a great deal of checking had to be done using our standard library reference sources and telephone directories. A number of afterburner operations have been reported in the literature, and these were added to our list as they were located. In addition, a number of questionnaires were sent to companies where we did not have specific information that they had afterburners in operation, but their plants were such that we would expect them to have such units (i.e., oil companies, chemical companies, coffee suppliers, etc., especially located in California).

A total of 559 addresses were obtained in this way, and most of them were known to have had afterburners in operation at one time.

20.3.5 Replies

Replies were received from 156 plant locations, of which 53 replied that they did not have afterburners. The remaining 103 had a total of 238 afterburners installed. Many of these replies were only obtained after repeated solicitations both by letter and by telephone calls.

The quality of the replies varied considerably depending on the sophistication of the plant engineering staff. In most cases test data had not been taken, and in others the companies were reluctant to divulge it. Purchase and installed costs were available in a large percentage of the cases. Many plants do not keep records of fuel costs for afterburners separately, so in these cases, operating costs were not available.

20.3.6 Treatment of Information

The information obtained from the questionnaires was tabulated in computer format (see Chapter 11, Table 11-2). This allowed internal manipulation of the data including sorting by category, adjusting purchase and installed costs to 1970 values, adjusting operating costs to a 24 hour basis, computation of average residence time and machine plotting selected data.

Process categories were selected using the Standard Industrial Classification Manual. The classification used was by the immediate process to which the afterburner was fitted, rather than the overall description of the manufacturing facility. For example, painting of automobiles and of metal cans were both included under 3479 (Coating), rather than using 3712 (Motor Vehicle Bodies) and 3411 (Metal Cans), respectively.

A listing of the process categories used is given in Table 20-1.

20.3.7 Follow-up Visits

Follow-up visits were made to users of afterburners who had specific information and wide experience and had no objections to a visit by our staff. Most companies were willing to cooperate but few had data or experience enough to justify a visit trip. Telephone conversations were sufficient in most cases. Visits were made to:

1. Automobile body painting operations in the Los Angeles area.
2. Can companies in Chicago and San Francisco with wide experience of both thermal and catalytic units.
3. A large chemical company in Philadelphia with experience on phthalic anhydride.
4. A coffee company which has recently replaced two old afterburners with a Proctor and Schwartz Traps Unit.
5. A number of smaller companies in the Bay Area on widely differing industries from asphalt blowing to drum manufacture.

Table 20-1. CLASSIFICATION OF USERS BY STANDARD INDUSTRIAL CLASSIFICATION MANUAL

8061	Hospitals
3900	Miscellaneous manufacturing industries
3714	Motor parts and accessories
3641	Electric lamps
3621	Electric motors and generators (rebuilding)
3479	Coating and engraving (enameling and painting)
3341	Secondary smelting and refining non-ferrous metals
3323	Steel foundries (investment casting)
3296	Mineral wool (glass, wool, fiberglass)
3292	Asbestos products (brake shoes vinyl tiles)
3079	Miscellaneous plastics products (e.g., phenolic laminates, film)
3069	Fabricated rubber products (foam rubber)
3000	Rubber and miscellaneous plastic products
2911	Petroleum refining (including asphalt)
2900	Petroleum refining and related industries
2851	Paints, varnishes, lacquers, enamels and allied products
2834	Pharmaceuticals
2815	Cyclic intermediates (phthalic anhydride, toluene, xylene, etc.)
2800	Chemicals and allied products
2700	Printing, publishing and allied industries
2426	Hardwood dimension and flooring mills
2095	Roasted coffee
2094	Animal and marine fats and oils (rendering)
2013	Sausages and other prepared meat products (smokehouses)
2000	Food and kindred products
4952	Sewage systems

20.4 Determination of Major Sources and Potential Applications

Emission inventory information was found to be quite variable in scope, availability and existence. Data is especially sparce relative to hydrocarbon emissions since previous concern has been with particulates and SO_2. Some states have developed complete information whereas others have none. The emission estimates that we used were primarily based on emission factors and total production of materials that normally find their way into the atmosphere (e.g., solvents). Projection of emissions in the future was then tied to the output growth of a particular industry or to the population growth. A tentative ranking of the emissions was made by combining quantity, toxicity and nuisance factors.

20.5 Research Recommendations

The research recommendations are based on the deficiencies identified in the review of technology. The deficiencies are discussed in the various sections in this handbook describing the technology. The research and development recommendations are presented in Chapter 22.

User's Survey Follows

2886

SHELL DEVELOPMENT COMPANY
A DIVISION OF SHELL OIL COMPANY
1400-53rd STREET
EMERYVILLE, CALIFORNIA 94608

April 8, 1971

User Survey - APCO Afterburner Systems Study
Contract EHS-D-71-3

We are currently engaged in an Air Pollution Control Office (APCO) sponsored engineering study of afterburner systems for control of gaseous or gas borne emissions from stationary sources. APCO was formerly the National Air Pollution Administration and is a part of the new Environmental Protection Agency. The objectives of this study are to increase the effectiveness and utilization of flame and catalytic afterburners. Upon completion of our work, we will publish an "afterburner handbook", to be available publicly, containing information on the following subjects:

1) What emissions are suitable for treatment by afterburning. Which components might cause difficulties.

2) What design features are desirable with regard to effectiveness, trouble-free operation, and low capital and operating costs.

3) What types of units are available. Who are the active manufacturers.

4) Unbiased data on initial costs and operating expenses including maintenance.

This handbook will be most useful in guiding the selection of required control equipment.

We believe that industrial users of afterburner equipment comprise an important source of information for this study, particularly concerning specific applications of afterburners as pollution abatement devices. Of special interest are the ability of these systems to perform satisfactorily in these applications; and costs associated with acquiring

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and operating afterburner systems. For this reason, we are surveying a representative sample of afterburner-users.

We ask that you participate in this survey by supplying the information requested on the enclosed questionnaire. Your participation will be valuable even if you can only supply part of the information requested. It is especially important to the success of the study that you point out any present problem areas so that possible improvements can be identified. The information you supply will be held confidential. Some of it may appear in tabular or statistical form in our reports and handbook but without identifying your company.

We would appreciate your filling out a separate copy of pages 2-6 for each afterburner application or each type of unit. We can supply additional copies if needed. A self-addressed, postage-paid, envelope is enclosed for the return of the completed questionnaire. If you have any questions, please call us collect at: (415) 653-2100 and ask for E. R. Slater, R. D. Hawthorn, R. W. Rolke, C. R. Garbett, or G. D. Towell.

Your cooperation in participating in this survey will be greatly appreciated.

Sincerely yours,

G. D. Towell

G. D. Towell,
Project Manager

GDT:psb

Enclosures: Survey Questionnaire
Return Envelope

288d
OMB-85-S70036
Expires 9/30/71

NAPCA CONTRACT EHS-D-71-3

Survey of Users of Afterburner Equipment

Date:________________

Section I. COMPANY IDENTIFICATION

1. Name and location of company:

   a. Name:__________________________________

   b. No., Street:______________________________

   c. City:______________ State:__________ Zip Code:_____

2. Location of plant if different from above:

   a. Plant/Division:____________________________

   b. No., Street:______________________________

   c. City:______________ State:__________ Zip Code:_____

3. Person to contact regarding information contained in this report:

   a. Name:__________________________________

   b. Department/Division:________________________

   c. Telephone: (Area Code)______________________

4. Principal products manufactured at this plant:________________

   ____

5. How many afterburner systems do you have at this location? (If two or more, please complete pages 2 to 6 for each system.) _____

6. Person completing questionnaire (if different from above):

   a. Name:__________________________________

   b. Department/Division:________________________

   c. Telephone: (Area Code)______________________

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Section II. GENERAL INFORMATION

1. a. Name of company completing this form: ____________________

   b. Afterburner identification (if more than one at location): ____________

   c. Date of installation: ____________________

   d. Name of manufacturer of afterburner: ____________________

   Did this same company design and install the system? ____________

   If not, name of company(s) which did: ____________________

   ____

2. Name of process generating waste gas stream (varnish cooker, paint spray booth, etc.): ____________________

3. Capacity of process (lb/hr, items/day, etc.): ____________________

4. Is the afterburner operated principally to control (check one):
   a. Odor nuisance
   b. Emission of hydrocarbons (lb/hr or ppm level at afterburner inlet)
   c. Toxic nuisance
   d. Smoke or other particulate emission
   e. Other (such as a condensing plume) - Please specify

5. a. Does the afterburner provide sufficient emission control? ____________
   If not, please specify (residual odor, residual hydrocarbon emission, etc.)

   b. What problems have you had with keeping the system operable?

   Approximate % of time afterburner is inoperable: ______%

Section III. WASTE GAS STREAM FED TO AFTERBURNER

1. Description of waste stream leaving process:

   a. Average flow rate: ____________ SCFM

   b. Average temperature: ____________ °F

   c. Average composition

   Oxygen: ____________ % Volume

   Combustible contaminant: { ____________ (ppm as hexane or other basis)
   ____________ lb/hr of contaminant

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d. Heating value of waste stream ____________________ BTU/SCF

e. Are the flow rate, composition, and temperature highly variable? (process operated batchwise, etc.) ______

If so, what are the ranges in flow rate? ______ to ______ scfm

composition? ______ to ______ ppm

temperature? ______ to ______ °F

f. Does the waste stream consist of anything other than gaseous hydrocarbons, air, water vapor, and CO_2? (chlorinated hydrocarbons, inorganic dust, liquid mist, etc.) ______

If yes, please specify____________________.

____

g. How many hours per day is process generating waste stream?______

2. Is the waste stream pretreated (other than heat exchange) prior to afterburning? (scrubbing, filtration, air dilution, etc.) ______

If yes, please specify____________________

____

3. Is a heat exchanger used to preheat the waste stream before afterburning?______

If yes, are the hot gases leaving the afterburner used as the heating medium? ______

Section IV. THERMAL AFTERBURNER UNITS (If unit is catalytic type, skip to Section V)

1. Operating temperature

Preheater (if used) inlet ______°F outlet ______°F

Afterburner inlet ______°F outlet ______°F

2. Operating pressure____________________(inches of water or psig)

3. Fuel type (check one):

a. Natural gas

b. Fuel oil

c. Other (please specify)____________________

4. Rate of fuel consumption ____________________

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5. Combustion chamber dimensions

Diameter____________ Length __________________

6. Is air supplied to the burner? (other than in waste stream)_________________

Section V. CATALYTIC AFTERBURNER UNITS (If unit is thermal type, skip to Section VI)

1. Operating temperature

Heat exchanger preheater (if used) inlet ______°F outlet ______°F
Afterburner inlet ______°F
After direct fired preheater ______°F
Outlet of catalyst bed_____________°F

2. Operating pressure________________________________(inches of water or psig)

3. Is a direct fired preheater used? ______

If yes, a. What is fuel type? (check one)

1) Natural gas

2) Fuel oil

3) Other (please specify) ____________________________________

b. Rate of fuel consumption ____________________________________

c. Is air supplied to the burner? (other than in waste stream)_____

4. Catalyst

a. What is catalyst configuration? (check one)

1) Mesh mats

2) Parallel cylinders

3) Pellets in baskets

4) Other (please specify) ____

b. What is active catalyst constituent? (copper oxide, platinum, etc.) _____

____

What is the nature of the catalyst carrier? (metal, ceramic, etc.) _____

____

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c. Typical life of catalyst before replacement ____________________ months

d. Is periodic cleaning of catalyst unit employed? ____________________

If yes, 1) How frequently? ____________________

2) What method is used? (air blasting, water wash, etc.) ______

____

Section VI. TYPICAL PERFORMANCE DATA

a. Inlet pollutant concentration (If analyze for individual compounds, please specify each) ____________________ ppm

____________________ ppm

b. Outlet pollutant concentration ____________________ ppm

____________________ ppm

c. Do you have specific test data? ______

Would you make it available to us for this study? ______

d. Do you monitor performance of the unit? ______ How? (check one)

1) Continuously ______ 2) Periodically ______ 3) Occasionally ______

Section VII. GASES LEAVING AFTERBURNER

1. Is a heat exchanger used to recover heat from gases leaving afterburner? ______
2. Is part of the hot afterburner effluent recycled to the process? ____________
3. Is posttreatment (other than heat exchange) employed? (scrubbing, filtration, etc.) ______

If yes, please specify____________________

____

Section VIII. OTHER SYSTEM COMPONENTS

1. What safety features are included in the design? (flame arrestor, high or low temperature shutdown, shut down if burner flame is lost, etc.)____________

____

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2. Is a separate fan used in the afterburner system? ____________________

   If so, is it located (check one)

   a. Before afterburner

   b. Following afterburner

3. Are unusual materials of construction (other than carbon steel and normal refractory) used? ______

   If yes, please specify ____

   ____

Section IX. COSTS

1. Approximate installed cost of afterburner system $____________________

2. Purchase cost of afterburner unit itself $____________________

3. Approximate operating cost of afterburner system ____________________ $/year

4. Do you have a more detailed breakdown of operating and capital costs which you would discuss with us? ______

Section X. FOLLOWUP

Would you be willing to discuss in more detail system performance, cost data, and design features through a telephone call or visit to your plant by one of our representatives? ______

If you have any questions, please call one of the people listed in the cover letter at (415) 653-2100. An envelope has been enclosed for return of this questionnaire to Shell Development Company, P. O. Box 24225, Oakland, California 94623, ATTN: G. D. Towell. Thank you again for your cooperation.

RWR:pm
Enc.

Chapter 21. MAJOR SOURCES AND POTENTIAL AFTERBURNER APPLICATIONS

21.1 Summary of Investigation Results

The rate at which organic pollutants are emitted in the United States from both mobile and stationary sources has been estimated for 1968 and 1970, and projected through 1980, 1990, and 2000. The results are summarized in Table 21.1. The contribution of stationary sources is projected to rise from 50% of the total in 1970 to 85% in 2000. The projections of emissions from stationary sources were made with the assumption that the present levels of legislated emissions control would not change during the projection period. The projected mobile source emissions reflect the stringent automobile emission regulations which are scheduled to take effect in 1975.

Several of the largest sources shown in Table 21.1 are either mobile or are not amenable to afterburner control. Examples are motor vehicles, refuse disposal, gasoline marketing, etc. Conversely, many sources which do not emit nationally significant tonnages of combustible emissions are, non-the-less, prime candidates for afterburner control because they constitute severe local nuisances. Odor and smoke sources, e.g. rendering plants, coffee roasters, brake shoe debonding ovens, etc., frequently fall into this category. No attempt was made to estimate the total rates of emissions of effluents which are objectionable primarily because of malodor. Instead, a count of the number of point odor sources was used to give a measure of the severity of the odor pollution problem. The results are shown in Table 21.2.

Emissions from sources amenable to afterburning have been characterized by the type of compound emitted. Each set of emissions has been assigned a ranking based upon toxicity, malodor, and photoreactivity to allow an importance comparison of the various pollutant sources. In addition, each source has been assigned a ranking based upon the relative ease of control by afterburning. The source characterization and rankings are given in Table 21.3.

Criteria to be considered when choosing between afterburners and other control devices for particular applications have been outlined. The costs associated with extensive afterburner application to control stationary sources emitting nationally significant amounts of combustible pollutants excluding, however, control of sources objectionable because of odor, is estimated to be $360 million for equipment purchase and installation. The fuel cost associated with this level of control (again excluding control of odorous sources) is estimated to be $466 million for the purchase of 0.931 trillion scf of natural gas annually. These estimates are based upon achieving the maximum feasible level of control which would reduce the total emissions shown in Table 21.1 for 1970 by 11.1 billion lbs per year, or about 15%. At this level of control, approximately half of the total emissions from the stationary sources suitable for afterburning would be incinerated.

It should be emphasized again that these costs make no provision for controlling sources objectionable primarily because of malodor or carbon monoxide, and they would be increased if these costs were included. The assumptions underlying these estimates are given in section 21.3.3.

Table 21.1. ESTIMATES OF NATIONWIDE EMISSIONS OF ORGANIC CHEMICALS FROM MOBILE AND STATIONARY SOURCES

	Emissions, 10^6 lbs				
	1968	1970	1980	1990	2000
Mobile Sources					
Motor Vehicles	36300	33200	8810	4320	4900
Other Transportation	2000	2030	3090	4700	7160
Total Mobile	38300	35200	11900	9020	12060
Incidental Sources					
Forest Fires	4400	4400	4400	4400	4400
Coal Banks	400	400	400	400	400
Building Fires	200	200	200	200	200
Total Incidental	5000	5000	5000	5000	5000
Stationary Sources					
Solvent Evaporation	13400	14500	22100	38100	75000
Refuse Disposal	3430	3510	880	1330	2010
Agricultural Burning	3400	3400	3400	3400	3400
Refineries	2940	3340	4300	5620	7370
Gasoline Marketing	1710	1850	2470	3440	4760
Carbon Black Manufacture	1570	1450	1390	1630	2150
Wood Burning	800	800	800	800	800
Charcoal Manufacture	417	426	649	989	1510
Metallurgical Coke Manufacture	389	404	404	404	404
Fuel Oil Burning	324	364	441	567	924
Petrochemical Manufacture	168	194	484	1200	3020
Automobile Body Incineration	6	6	9	12	13
Smoke Houses	1	1	1	1	2
Coffee Roasting	1	1	2	2	2
Steel Drum Reclaimation	1	1	1	1	1
Brake Shoe Debonding	< 1	< 1	< 1	< 1	< 1
Total Stationary	28600	30200	37300	57500	101400
Total All Sources	71900	70400	54200	71500	118500

Table 21.2. THE TOTAL NUMBER OF POINT ODOR SOURCES[a)]

Odor Source	No. of Plants
Plastics Fabrication	5000
Meat Packing	2700
Pharmaceutical Manufacturing	875
Poultry Dressing	840
Soap Manufacture	668
Fish Canning	320
Rendering	268
Breweries	185

a) *Census of Manufactures*, 1967.

21.2 Applicability of Afterburners to Air Pollution Control Problems

Incineration can control the emissions of any gases, vapors, or aerosols which are combustible or thermally decompose at high temperatures. Most applications are for the control of effluent streams objectionable because of malodor or smog potential, e.g., rendering or paint baking effluents, but incineration has also been applied successfully to effluents containing combustible particulates, e.g., phthalic anhydride manufacture and smokehouse effluents. However, when an effluent stream is objectionable because of particulates content only, mechanical collection devices such as fabric filters or wet collectors give highly efficient control at a considerable cost saving relative to afterburners.[21-19]

Before incineration is chosen as a control technique, it should be determined that the combustion products are not offensive pollutants themselves. For example, incineration of organic chlorides produces hydrogen chloride, a serious pollutant. Other effluent components producing undesirable combustion products are the other halide compounds, nitrogenous compounds or sulfurous compounds. When incineration is the only feasible control method for a particular effluent it is possible, of course, to cleanse the afterburner effluent using a secondary control device, e.g., a wet scrubber. In this way the range of incineration applicability is considerably extended.

Finally, incineration requires that the effluent be collected and ducted to the afterburner with minimal addition of diluent air. The fuel costs for afterburning are directly proportional to the total volume of effluent to be treated. Thus unless the effluent can be efficiently collected the cost of afterburning may be prohibitive.

Malodorous emissions are particularly well suited to afterburner control. Alternative methods such as scrubbing or adsorption frequently cannot achieve the high degree of removal of malodorous gases and vapors required for satisfactory odor control. This is not a problem for well designed afterburners, however. In addition, the afterburner can control combustible particulates which sometimes accompany malodorous gases.

21.3 Combustible Air Pollutants: The Nature and Dimensions of the Problem

We have estimated the rate at which combustible compounds are emitted in the United States for 1968 and 1970 from processes which are and are not amenable to afterburner control. The emission rates have been projected thru the years 1980, 1990 and 2000, thus giving an indication of how the pollution picture will change in the next 30 years if further controls are not implemented. In most cases it was assumed that either there were no controls at present, or that only those portions of an industry located in California were presently controlled. Solvent emission sources are the incinerable sources most significant on a national basis. As shown in Tables 21.1 and 21.3 the total estimated solvent emissions for 1970 are 14.5 billion lbs; only the gasoline powered automobile, emitting 32 billion lbs, accounts for a greater tonnage. Other significant sources of incinerable emissions are refuse incineration, oil refineries, gasoline marketing, carbon black manufacture, charcoal manufacture, metallurgical coke manufacture, and chemical plants (as represented by the petrochemical portion of the industry).

Table 21.3 follows

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Table 21-3. CHARACTERIZATION OF COMBUSTIBLE EMISSION SOURCES

SOURCE \ CHARACTERIZATION	No. Sources or approx. 1970 Emission, MM lbs of Organics c)	Ease of Afterburner Control d)	KNOWN EMISSIONS													Index of Emission's Offensiveness e)
			Hydrocarbons		Organic Halides	Organic Acids	Alcohols	Aldehydes	Nitrogenous Organics	Esters	Ketones	Ethers	Sulfurous Organic or H_2S	Odor	Other	
			Aromatic	non-aromatic												
SOLVENT USAGE	1400 MM lbs		X	X	X		X	X	X	X	X	X		X		
adhesives		2	X	X			X			X	X					3.7
degreasing metal parts		1			X											6.2
dry cleaning		2	X	X	X									X		4.1
fiberglass curing		3		X										X		2.7
graphic arts		3	X	X			X			X	X	X		X		3.6
herbicides		1	X	X												3.5
rubber products fabrication		3	X	X	X						X			X		3.5
surface coating manufacture		3	X	X			X			X	X	X		X		3.6
surface coatings (industrial)		3	X	X			X		X	X	X	X				4.3
surface coatings (trade)		1	X	X			X			X	X	X				3.6
surface coatings removers		1	X	X	X					X						2.8
FOOD AND BYPRODUCTS				X		X		X	X				X	X		
breweries	175 plants	3												X		-
charbroiling restaurant	a)	3												X	smoke	-
coffee roasting	1.5 MM lbs	3	X	X		X		X							smoke	4.6
deep fat frying	a)	3		X											smoke	2.7
fish canning	320 plants	2							X					X		8.5
meat and bone meal drying	a)	1												X		-
meat packing plants	2700 plants	1												X		-
poultry dressing plants	840 plants	2												X		-
rendering plants	265 plants	2							X					X	ammonia	8.4
smokehouses	1.2 MM lbs	3	X	X		X		X			X					4.5
vegetable oil extraction	358 plants	3		X												2.7
COMBUSTION PROCESSES				X		X		X						X	smoke	
auto body incineration	5 MM lbs	3		X		X		X							smoke	4.7
brake shoe reclamation	0.2 MM lbs	3		X		X		X								4.7
copper wire insulation burnoff	<2. MM lbs	3		X		X		X								4.7
fuel oil burning	564 MM lbs	b)		X				X							SO_2	4.5
refuse incineration	5985 MM lbs	3		X		X		X						X	smoke	4.7
secondary nonferrous metal recovery	405 plants	3													smoke	-
steel drum reconditioning	0.8 MM lbs	3		X		X		X							smoke	4.7
GENERAL INDUSTRIAL PROCESSES			X	X	X	X	X	X	X	X	X	X	X	X		
carbon black manufacture	1449 MM lbs	3		X												2.7
charcoal manufacture	426 MM lbs	3	X	X		X	X								combustible particulates	4.0
chemical plants	194 MM lbs	1	X	X	X	X	X	X	X	X	X	X	X	X		4.9
hardboard manufacture	a)	3											X			2.7
gray iron foundries	1061 plants	2													CO	-
metallurgical coke manufacture	289 MM lbs	2	X	X				X						X	ammonia	5.5
oil refineries	2058 MM lbs	2	X	X									X	X		4.4
paper manufacture (pulp mills)	279 plants	1											X	X		8.5
pharmaceutical manufacture	875 plants	3												X		-
plastics fabrication	5000 plants	1	X	X										X		5.5
soap manufacture	668 plants	3												X		-
MISCELLANEOUS SOURCES			X	X		X		X						X		
automobiles	52060 MM lbs	b)	X	X		X		X							CO, NO_x	4.6
gasoline marketing	1552 MM lbs	b)	X	X												3.5
oil tempering of metals	a)			X											smoke	2.7
sewerage systems	a)													X		-

a) Information not available.
b) Emissions have been included even though the process is not amenable to afterburner control.
c) MM is used to designate million in tables throughout this chapter.
d) A rank of 1, 2, or 3 was assigned to each source according to whether it was easily controllable, was controllable with considerable difficulty, or control was impossible or possible only with extreme difficulty, respectively.
e) The index can range from zero for emissions with no offensive qualities to ten for an emission with maximum offensiveness.

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Not all of these sources are controllable by afterburners, e.g., automobiles, open burning of refuse, some of the industrial process emissions, and some of the solvent emissions, but they have been included to give a clearer picture of the total problem.

Of the large tonnage emission sources shown in Table 21-1, afterburning is commonly applied only to solvent sources. In the Los Angeles and the San Francisco Bay areas afterburners have played an important role in the efforts to control photochemically reactive emissions. As photochemical smog becomes a problem in other metropolitan areas, afterburners will be applied to this problem on a much wider scale.

Many smoke and odor sources, while not significant in terms of national tonnage, do represent severe local nuisances. These sources are the most common applications of afterburner control. Typical examples are emission control for smokehouses, coffee roasters, rubber curing ovens, waste incinerators, and rendering processes. The offensiveness of odorous emissions is only weakly reflected in the total tonnage emissions, because odor intensity is proportional to the logarithm of the odorant concentration, as described by the Weber-Fechner law,

$$I = k \ln C,$$

where I is the intensity of odor sensation,
k is the constant peculiar to the odorant in question, and
C is the odorant concentration.

For any concentration above the perception threshold, the apparent intensity increases slowly with increasing concentration. This implies 1) that a large scale odor source may not be a much greater nuisance than a small source, and 2) that an odor control procedure must remove a very high percentage of the odorant to be effective.

21.3.1 Types of Emissions Not Covered in This Study

No attempt was made to estimate the rate of emissions from sources which are offensive because of severe odor problem. For these cases, it was felt that the total number of point sources would give a better estimate of the magnitude of the pollution problem. The number of sources was also given for emission sources for which there was no basis available to compute the rate of emissions. These counts are shown in Tables 21.2 and 21.3.

Combustible particulate emissions were excluded from the emission rate estimates. There were two primary reasons for this exclusion. First, published data on particulate emissions rarely differentiate between combustible organic particulates emitted by a process and noncombustible inorganic particulates which may be generated by a process. Second, as mentioned above, incineration is an uneconomic method for particulates removal.

Also, carbon monoxide (CO) was excluded from the estimates because industrial stationary sources contribute less than ten percent of the nationwide emissions.21-31) Thus maximum possible contribution to lowering ambient CO levels by afterburning is only about 10 percent.

21.3.2 Emission Characterization

The chemical compounds known to be present in emissions from various sources are identified in Table 21.3. Each emission has been assigned an arbitrary index to characterize the relative offensiveness of the emission. The index was based upon the toxicity, odor potential and photochemical reactivity of the emission's components. This index is one of many possible indices, and another investigator would likely derive a different index from the same facts. The rankings should be used with this in mind. The index has a range of from zero for no objectionable qualities to 10 for the maximum offensiveness. Assigning an overall ranking to each source was considered, but rejected because an equitable basis for such a ranking was not found.

It is hoped that the information presented will be sufficient to allow interested users to utilize it in devising source ranking schemes which reflect their particular requirements. This is, in fact, the primary reason for conducting this phase of the study. A detailed presentation of the emission ranking system is presented in 21.4. The methods used in projecting levels of pollutant generating activity and the methods used in estimating emissions are given in detail in 21.5 and 21.6 respectively.

Each source was also considered from the viewpoint of the suitability of applying afterburners to control their combustible emissions. A rank of 3, 2, or 1 was assigned to each source according to whether it was easily controllable, was controllable with considerable difficulty, or control was impossible or possible only with extreme difficulty, respectively. For example, degreasing was ranked one because the trichloroethylene and perchloroethylene used as solvents are very difficult to burn, and yield noxious combustion products, while fiberglass curing was ranked three because the fumes are easily collected, burn well, and yield carbon dioxide and water as combustion products.

21.3.3 Potential Emissions Reduction Through Extensive Afterburning

If nationwide pollution control laws required the maximum feasible application of afterburning to stationary sources, it is possible that in 1970 nationwide emissions could have been reduced by about 11.1 billion lbs. This figure is a rough estimate which was obtained by assuming that it is possible to control fifty percent of the organic solvent emissions and organic emissions from metallurgical coke manufacture, oil refineries, and chemical plants, and that the emissions from all sources in Table 21.3 with ease of control rank 3 could be controlled at the 95 percent level. It is instructive to consider a rough estimate of the costs associated with this level of control. We assume that the organic emissions are in a 400°F fume at a concentration of 0.1% v as hexane, and that they will be incinerated at 1400°F using natural gas as fuel with no heat recovery. At 0.1% v hexane, there are 1000 scf fume/scf hexane,

$$\text{scf fume} = \left(\frac{1.10 \quad 10^{10} \text{ lb hexane}}{86 \text{ lb/lb mole hexane}}\right)\left(\frac{379 \text{ scf hexane}}{\text{lb mole hexane}}\right)\left(\frac{1000 \text{ scf fume}}{\text{scf hexane}}\right) =$$

$$4.85 \times 10^{13} \text{ scf fume/year}$$

From Figure 11-10, operation of these afterburners at 1400°F will require the burning of 0.0192 scf natural gas/scf fume.

required gas = $(4.85 \times 10^{13}$ scf fume)(0.0192 scf gas/scf fume) = 0.931 trillion scf gas/year.

gas cost at $\$.50/10^{3}$ scf = \$466 million/year

The fume flow of 4.85×10^{13} scf/year corresponds to a flow rate of about 9×10^{7} scfm, assuming 24 hour per day, 7 days per week operation. If we assume a purchase cost of \$2/scfm (a low estimate as per figure 11.1 or 11.2) the purchase cost of the required afterburners would be \$180 million. Installed costs would be about twice this amount, or \$360 million. This cost would be quadrupled by estimating on the five day - 40 hour week basis which is common to many of the industries emitting combustible pollutants. Heat recovery could reduce the fuel usage by about half, but would increase the purchase costs by a factor of three or four. For comparison, the marketed production of natural gas in the United States in 1970 was 21.9 trillion scf. It should also be noted that the calculations make no provision for controlling malodorous fumes whose emission rates were not estimated in this study, or for controlling CO emissions. The very high cost and natural gas consumption associated with extensive afterburning should be considered when pollution control policies are formulated by regulatory agencies.

21.4 Emission Ranking Procedure

The objectionable factors which were used to characterize the degree of offensiveness of a particular compound were toxicity, odor, and photochemical reactivity. Toxicity factors were based on the threshold limit values (TLV) for airborne contaminants adopted by the American Conference of Governmental Industrial Hygienists (ACGIH-.[21-7]) The TLVs are given in the second column of Table 21-4. A toxicity factor (TF) was then assigned according to the scheme shown in Table 21-5. For compounds which have not been assigned TLVs by the ACGIH, a TF was estimated by interpolating between TLVs for compounds of similar structure. For each class of compounds the TFs were then averaged to obtain a suitable TF for ranking the entire class. The TFs are shown in column 3 of Table 21-4.

The odorant rankings were adapted from rankings proposed in a NAPCA sponsored study of the social and economic impact of odors.[21-8]) The original rankings were extended by assigning an odor factor (OF) of two to aromatic hydrocarbons and to organic halides, two classes of compounds which are included in the present study but which were not treated separately in the odor study. The OFs are shown in column 4 of Table 21-4.

Photochemical reactivities for the compounds were ranked using an adaption of the system proposed by Altshuller.[21-9]) Rankings for compounds not included in Altshuller's original work were obtained by comparing his rankings with the reactivities reported by other workers in the field.[21-10,11,12,20,21]) The comparisons were rather arbitrary since different investigators used different measures of photochemical reactivity, and also because data could not be found for all compounds of interest. Again the rankings for individual compounds were combined to form an average ranking for each compound class. The rankings, designated as smog factors (SF), are shown in column 5 of Table 21-4.

Table 21-5. TOXICITY FACTORS AS FUNCTIONS OF TLVS

TLV mg/m^3	TF
0 - 10	5
11 - 50	4
51 - 200	3
201 - 500	2
501 - 2000	1
2001 +	0

In order to obtain an overall index for the offensiveness of each effluent, these rankings were applied in the following manner. The TF and OF rankings were scaled to a 0 - 10 range by multiplying by 2 and 2.5 respectively. The overall index was taken as a weighted sum of the three factors. Weights of 3, 2, and 1 were assigned to the toxicity, odor, and smog factors respectively. Thus for an emission containing classes of compounds $C_1, C_2, \ldots C_n$, the overall index was computed by

$$\text{Index} = \frac{1}{6n} \sum_{i=1}^{n} (3)(2)(TF_i) + (2)(2.5)(OF_i) + SF_i$$

Consider a rendering plant, for example. The effluent contains nitrogeneous and sulfurous organics (Table 21.3), thus $n = 2$, and the offensiveness index is:

$$\text{Index} = \frac{1}{(6)(2)}[6(5) + 5(4) + 1 + 6(5) + 5(4) + 0] = 8.4$$

This formula was arbitrarily chosen by the author, and it is recognized that other combinations are possible. Note that this system is intended to provide a combined offensiveness index for all the compounds in an effluent, not an index for the various classes of compounds as entities. Some pollutants exhibit offensive qualities which were not included in the present rankings, e.g., particulate formation, allergenic potential, or phytotoxicity. Particulates were not included in the rankings because particulates were not included in the emission estimates for the reasons noted in 21.3.1.

Allergenic potential was not included because, apparently, nearly all aeroallergens are of natural origin, and thus not suitable to afterburner control.$^{21-13)}$ Although some industrial products, e.g., insecticides, linseed oil, and some organic solvents, are allergens, their contribution to the overall problem appears minimal. Phytotoxicity was not included in the rankings because apparently, among the common combustible pollutants, only ethylene is a significant phytotoxicant.$^{21-14)}$ (Products produced in smog reactions do show phytotoxicity, however.)

21.5 Pollution Generating Activity Estimates and Projections

Three methods were used to project the levels of air pollutant generating activities. The first method tied future activity levels to the rate of increase of the population as projected by the U.S. Bureau of the Census series D projection.$^{21-3)}$ This method was applied to projecting levels of coffee roasting and meat smoking, using base data from the Census of Manufactures, 1967.$^{21-15)}$

Tables 21-4 and 21-6 follow

Table 21-4. TOXICITY, SMOG, AND ODOR FACTORS FOR COMBUSTIBLE POLLUTANTS

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	TLV[a] mg/m^3	Toxicity[c] Factor	Odor[d] Factor	Smog[e] Factor
Aldehydes		3	3	5
acetaldehyde	360	2		
acrolein	0.25	5		
butyraldehyde	b)	2		
formaldehyde	6	5		
Nitrogen-Organics		5	4	1
acrylonitrile	45	4		
aniline	19	4		
methyl amine	12	4		
toluenediisocyanate		5		
nitrobenzene	5	5		
Esters		2	2	1.5
acrylates	35-100	4		
amyl acetate	525	1		
n-butyl acetate	710	1		1
other butyl acetates	950	1		2
ethyl acetate	1400	1		1
ethyl hexyl acetate		1		
isobutyl isobutyrate		1		
isopropyl acetate	950	1		2
methyl acetate	610	1		
Ethers		1	1	5
ethyl ether	1200	1		6
glycol ethers	80-740	2		3
isopropyl ether	2100	0		[illegible]
Ketones		2	2	3
acetone	2400	0		0
cyclohexanone	200	3		2
diacetone alcohol	240	2		3
diisobutyl ketone	290	2		3
isophorone	140	3		3
methyl ethyl ketone	590	1		1
methyl isobutyl ketone	410	2		4
Non-Aromatic Hydrocarbons		1	1	5
acetylene	A	0		0
butadiene	2200	2		8
cyclohexane	1050	1		1
ethylene	A	0		4
higher olefins	1000	1		7
napthas	400	2		5
propylene	A	0		7
turpentine	560	1		5
Aromatic Hydrocarbons		2	2	4
benzene	80	3		0
ethylbenzene	435	2		3
styrene	420	2		8
toluene	750	1		3
xylenes	435	2		8
Organic Halides		2	2	2
allyl chloride	3	5		5
benzyl chloride	5	5		3
o-dichlorobenzene	300	2		0
p-dichlorobenzene	450	2		0
ethyl chloride	2600	0		0
methyl chloride	210	2		0
methyl chloroform	1900	1		0
methylene chloride	1740	1		0
perchloroethylene	670	1		0
trichlorethylene	535	1		5
vinyl chloride	1300	1		4
Organic Acids and Anhydrides		4	3	0
acetic acid	25	4		
cresol	22	4		
maleic anhydride	1	5		
phenol	19	4		
xylene based acids (e.g., phthalic anhydride)	12	4		
Alcohols		2	1	2
amyl alcohol		2		4
n-butyl alcohol	300	2		3
sec-butyl alcohol	450	2		3
tert-butyl alcohol	300	2		0
decyl alcohols		2		
ethyl alcohol	1900	1		2
glycerine (mist)	N	0		
iso-octyl alcohol		3		
isopropyl alcohol	980	1		0
methyl alcohol	260	2		
methyl isobutyl carbinol	100	3		
oxo-alcohols				
Sulfurous Organics and H_2S		5	4	0
dimethyl sulfate	5	5		
ethyl mercaptan	1	5		
hydrogen sulfide	15	4		
methyl mercaptan	1	5		

a) Threshold Limit Values of Airborne Contaminants, adopted by American Conference of Governmental Industrial Hygienists, 1970. "Threshold limit values (TLV's) refer to airborne concentrations of substances and represent conditions under which it is believed that nearly all workers may be repeatedly exposed day after day without adverse effect". A and N in place of a TLV refer to simple asphyxiants and inert nuisance dusts, respectively.

b) Blank spaces in the table indicate that information was not available.

c) Toxicity factors range from zero to five for minimum to maximum toxicity respectively.

d) Odor factors range from zero to 4 for none to maximum odor offensiveness.

e) Smog factors range from zero to 10 for none to maximum smog precursor activity.

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Table 21-6. SOLVENTS PRODUCTION AND EMISSIONS

Solvent	No. of Base Years	% Growth Projected	Production Estimate					% Used as Solvent	Solvent Use, MM lbs.				
			1968	1970	1980	1990	2000		1968	1970	1980	1990	2000
HYDROCARBONS													
benzene	11	9	7,320	8,530	22,700	55,600	135,000	1.5	110	128	341	834	2,025
ethylbenzene	5	6	4,500	5,600	10,000	18,500	36,000	0.25	11	14	25	46	90
special napthas	5	1.2	7,080	7,730	8,650	9,700	11,000	100	7,080	7,730	8,650	9,700	11,000
toluene	17	9	5,050	5,030	13,800	36,300	94,500	15	758	775	2,070	5,445	14,175
turpentine[a]	18	-6	2.3	1.8	1.0	0.6	0.3	100	2.3	1.8	1.0	0.6	0.3
xylenes	9	5	3,867	3,150	5,100	8,200	1_,500	15	580	473	765	1,230	2,025
							Hydrocarbon Totals		8,541	9,101	11,852	17,256	29,315
ORGANIC CHLORIDES													
o-dichlorobenzene	25	7	61	65	123	245	470	30	18	20	37	74	141
p-dichlorobenzene	20	0.5	70	67	70	77	82	90	63	60	63	69	74
methyl chloroform	2	9	300	355	870	2,100	5,200	84	252	298	731	1,764	4,368
methylene chloride	20	10	302	418	1,180	3,500	10,900	60	181	251	708	2,100	6,540
perchloroethylene	20	9	637	702	1,800	4,600	12,000	80	510	562	1,440	3,680	9,600
trichloroethylene	20	4.5	519	610	920	1,500	2,400	98	498	598	902	1,470	2,352
							Organic Chloride Totals		1,522	1,789	3,881	9,157	23,075
ALCOHOLS													
amyl alcohol	GNP	4.3	38	44	63	104	163	47	18	21	30	49	77
n-butyl alcohol	19	5	433	500	900	1,600	3,000	20	87	100	180	320	600
other butyl alcohols	17	5	500	560	1,000	1,750	3,200	10	50	56	100	175	320
decyl alcohols	4	10	153	180	530	1,600	4,600	15	23	27	80	240	690
ethyl alcohol[a]	15	5	732	760	1,260	2,120	3,440	100	732	760	1,260	2,120	3,440
glycerine	7	2	100	105	128	156	190	100	100	105	128	156	190
isopropyl alcohol	–[b]	0 - 4	1,890	1,990	2,300	3,400	4,900	11	208	219	253	374	539
methyl alcohol	20	7	3,817	4,944	9,500	20,000	44,000	10	382	494	950	2,000	4,400
methyl isobutyl carbinol	8	2.5	43	46	58	74	92	83	36	38	48	61	76
oxo-alcohols	6	3	620	690	880	1,200	1,620	3	19	21	26	36	49
cresols	20	5	81	93	155	260	450	10	8	9	16	26	45
							Alcohol Totals		1,663	1,850	3,071	5,557	10,426
ESTERS													
amyl acetate	GNP	4	10	11	16	24	35	85	9	9	14	20	30
n-butyl acetate	13	1.5	63	72	85	100	115	83	52	60	71	83	95
other butyl acetates	8	8	60	64	136	290	620	10	6	6	14	29	62
ethyl acetate	20	4	179	159	240	370	570	90	161	143	216	330	513
2-ethyl-hexyl acetate	GNP	4	3	3	4	6	9	80	2	2	3	5	7
isobutyl isobutyrate	GNP	4	5	6	9	13	19	100	5	6	9	13	19
isopropyl acetate	5	4	48	52	79	120	180	75	36	39	59	90	135
methyl acetate	4	0	9	9	9	9	9	80	7	7	7	7	7
							Ester Totals		278	272	393	577	868
KETONES													
acetone	20	6.5	1,518	1,598	3,100	6,000	12,000	31	471	495	961	1,860	3,720
cyclohexanone	8	9	481	500	1,300	3,300	8,300	4	19	20	52	132	332
diacetone alcohol	GNP	4	87	95	145	240	330	70	61	67	102	168	231
diisobutyl ketone	GNP	4	86	95	145	240	330	63	54	60	91	151	208
isophorone	GNP	4	28	31	46	70	108	100	28	31	46	70	108
methyl ethyl ketone	19	7	451	500	1,000	1,950	3,900	74	334	370	740	1,443	2,886
methyl isobutyl ketone	7	4	182	210	300	430	620	100	182	210	300	430	620
							Ketone Totals		1,149	1,253	2,292	4,254	8,105
ETHERS													
ethyl ether	19	2.5	100	103	135	180	240	31	31	32	42	56	74
glycol ethers	3	8	441	520	1,200	2,800	6,600	40	176	208	480	1,120	2,640
isopropyl ether	6	10	6	7	21	62	180	60	4	4	13	37	108
							Ether Totals		211	244	535	1,213	2,822
MISCELLANEOUS													
nitrobenzene	6	10	398	500	1,500	4,300	12,000	3	12	15	45	129	360
					All Solvents Total				13,376	14,524	22,069	38,143	74,971

a) Amount used as solvent

b) IPA growth expected to go from zero to -4% in 1975

c) Production histories were obtained primarily from Chemical Economics Handbook by Stanford Research Institute and from Synthetic Organic Chemicals published by the U.S. Tariff Commission. Some data were obtained from the petrochemical series by F. G. Sullivan in Hydrocarbon Processing and from Solvent Emission Control Laws and the Coatings and Solvents Industry by the Environmental Sciences Services Corp., Stamford, Connecticut. Solvent usage fractions were obtained from the above named publications and also from Oil, Paint, and Drug Reporter Chemical Profile.

The growth in production of most industrial products was assumed to follow an exponential growth law. Examination of production data for commodities plotted against time on a semi-logarithmic plot reveals that exponential growth often persists over very long periods before a decay develops in the rate of growth. Because a decay inevitably develops in the growth at some time, however, a conservative approach was taken in fitting a straight line to the production data curves. In particular, when the latest data indicated a downward trend, heavy emphasis was given to the latest data in projecting future production rates. The additional constraint was imposed that growth rates were limited to a maximum of 10%/year. While this method can lead to considerable error over a period of thirty years, better estimates would have to be based upon a careful analysis of all product end use markets. Such an analysis was clearly beyond the scope of the present study.

The chemical industry was treated in a special manner. Because of the general complexity of the industry with an extremeley diverse mix of products and feedstocks, we confined attention to only a limited portion of the industry. Specifically the fraction of the industry processing the major petrochemical feedstocks, acetylene, benzene, propylene, xylenes, toluene, ethylene, butylenes, and methanol to their primary first generation derivatives was considered. It was felt that a treatment of this portion of the industry would give an estimate for the order of magnitude of the chemical industry's air pollution problems. For the larger volume feedstocks, the growth rate of the products were each evaluated separately, and the total consumption of feedstock was obtained by summation. For the butylenes, methanol, toluene, xylenes, and acetylene, the feedstock growth was estimated from past history, and the most recent product mix information available was assumed to apply for the period of estimation.

The third method used for predicting production rates was to assume that growth proceeded at a rate equal to that projected for the GNP. This method was applied only to products where insufficient historical production data was available. The second method was preferred wherever production data for three or more years could be found.

Solvent usage projections were made by multiplying the production estimate for each particular solvent by the fraction of total production which is used as solvent. Data on solvent usage was taken from *Chemical Profiles*, the *Chemical Economics Handbook*, and a study of the impact of solvent emission control laws.[21-16-17-18] The estimates of solvent production and usage are shown in Table 21-6. Also shown are the number of years for which past production data could be obtained and the projected annual production growth rate.

21.6 Emission Estimates and Projections: Methods and Results

We have estimated the emission rates of combustible pollutants for most of the large volume emissions sources. The methods used and detailed projections of the emission estimates are presented below for each source of effluent. As noted above, the effluent rates of sources objectionable primarily because of malodorous emissions have not been estimated. The data for the number of process point sources reported in Table 21-2 were taken from the *1967 Census of Manufactures*.

21.6.1 Solvent Emission Rates

Estimates of solvent emissions have been made on the assumption that all chemical and petroleum products sold as solvents will eventually find their way into the atmosphere. The fractions of production of solvent chemicals actually used as solvents were obtained from References 21-16, 17, and 18. It was also assumed that the fractions used as solvent remained constant through the 30 year projection period. The solvent emission estimates are shown in Table 21-6. Solvent end use markets are extremely diverse. Probably a large fraction of the solvents used commercially could be collected and incinerated if necessary, while those sold on the retail markets, e.g., in paints, polishes, etc., could not.

21.6.2 Oil Refinery Emissions

Oil refining is a very large industry in the United States, and refineries are a significant source of combustible emissions. As a result, an unusually large amount of information is available about the emission characteristics of refineries. This allows a detailed estimate of refinery emissions by source. The detailed emission estimates and projections are shown in Table 21-7 and summarized in Table 21-8. The emission factors from Reference 21-1 were used. It was generally assumed that only the 13% of the refining capacity located in California was controlled.[21-24] This assumption was made because controlling hydrocarbon emissions is generally not profitable and at the present time hydrocarbon emissions are specifically controlled only in California. Catalytic cracker emissions are an exception to this rule because the installation of a CO (carbon monoxide) boiler to utilize the heat content of catalyst regenerator off gases is an economically attractive investment. Consequently, it was assumed that 50% of present catalytic cracker capacity and all new capacity additions are controlled. The fractions of crude input which are fed to catalytic crackers and to vacuum distillation are assumed constant at 1971 values, 45% and 37% respectively.[21-24] The ratio of effluent water to crude capacity was taken as 1.11 bbl waste water/bbl crude capacity.[21-23] The cooling water circulation rate used was 41.3 bbl water/bbl crude, and the heat requirements were 1.94 x 10^5 Btu/bbl crude from refinery gas and 4.64 x 10^5 Btu/bbl crude for fuel oil.[21-25] Storage tank emissions were conservatively estimated by using the ratio of storage tank emissions to crude capacity reported for Los Angeles County, California.[21-23]

Possibly from 35 to 65% of the 1970 combustible pollutants from refineries could be controlled by extensive afterburning if required. The primary obstacle to control of many of these sources is the expense of collecting the emissions for incineration. Sources such as cooling towers, valve and flange leaks, and pump seals are too widely distributed for efficient control.

21.6.3 Gasoline Marketing Emissions

An average emission factor, 0.882 lb HC emitted/bbl gasoline, was used to estimate emissions from filling service station's and automobile tanks.[21-1] The projected gasoline production and marketing emissions are shown in Table 21-9. Incineration is not an efficient method for controlling emissions from this source; vapor recovery should be employed.

Table 21.7 REFINERY EMISSION BY SOURCE, MM LBS

	1968	1970	1980	1990	2000
Production est., MM bbl/yr	3774	4284	5600	7400	9800
cat crackers HC[a]	172.1	195.7	195.7	195.7	195.7
blowdown systems	988.9	1122.4	1467.2	1938.8	2567.6
process drains	769.9	873.9	1142.4	1509.6	1999.2
vacuum jets	157.8	179.1	234.1	309.3	409.6
cooling towers	39.2	44.6	58.2	77.0	101.9
pipeline valves and flanges	105.7	120.0	156.8	207.2	277.4
vessel relief valves	41.5	47.1	61.6	81.4	107.8
pump seals	64.2	72.8	95.2	125.8	166.6
compressor seals	18.9	21.4	28.0	37.0	49.0
other	37.7	42.8	56.0	74.0	98.0
storage tanks	462.8	525.4	686.7	907.4	1201.8
furnaces HC	60.4	68.5	89.6	118.4	156.8
HC, Total	2919.0	3313.0	4272.0	5581.0	7332.0
cat crackers - aldehydes	15.3	17.5	17.5	17.5	17.5
furnaces, aldehydes	9.2	10.5	13.7	18.1	23.9
Aldehyde, total	24.5	28.0	31.2	35.6	41.4

a) Assume 50% of 1970 capacity and 100% of post 1970 capacity increases are controlled by CO boilers, with zero emissions.

Table 21.8 REFINERY EMISSION, MM LBS

Year	Crude Oil Demand, MM bbl	HC	Aldehydes
1968	3,774	2,919	24.4
70	4,284	3,313	28.0
80	5,600	4,272	31.2
90	7,400	5,581	35.6
2000	9,800	7,332	41.4

Table 21.9 GASOLINE MARKETING EMISSION, MM LBS[a)]

Year	Gasoline Production, MM bbl	HC
1968	1,934	1,705
70	2,100	1,852
80	2,800	2,470
90	3,900	3,440
2000	5,400	4,763

a) Emissions from filling sertice station and automobile tanks.

21.6.4 Emissions from Refuse Burning

Data on the volume of solid wastes and methods of disposal were taken from Technical-Economic Study of Solid Waste Disposal Needs and Practices.[21-26] The assumptions used were:

1) that municipal waste incineration will increase as projected in the solid waste study;

2) that industrial waste incineration would increase proportionately with the GNP, i.e., at 4.3%/year;

3) that open dump burning will be eliminated by 1980;

4) that sawmill waste generation will level off after 1975.

The projected emissions are shown in Table 21-10. Note the sharp decrease in emissions forecast for 1980 which results from stopping open burning. Of the total emissions from refuse disposal only approximately 20 percent of the 1970 emissions which are from municipal and industrial incinerators (excluding "TeePee" burners) could be afterburned. This fraction could probably be controlled more efficiently using modern multiple chamber incinerators.

21.6.5 Emissions from Carbon Black Manufacture

The manufacture of carbon black is a large source of combustible emissions, as shown in Table 21-11. The decrease in emissions in 1980 results from the trend to replace natural gas fueled channel and furnace production with production from the more economical and more efficient oil fueled furnace process. Emission factors were taken from Reference 21-1. Possibly from 40 to 70% of the organic emissions from carbon black manufacture could be incinerated.

21.6.6 Emissions from Petrochemical Manufacture

There has been very little emissions data reported for chemical processes. However, because chemical plants are generally considered to be fairly large scale sources, it was decided to make a rough estimate of their contribution to combustible air pollution. It was assumed that chemical plants would emit at the same rate as an uncontrolled refinery, about 0.005 lb effluent/lb feedstock.[21-23] Further, half of the effluent was assumed to be emitted as the feedstock, and half was assumed to be converted to the reaction product. For example, assuming theoretical yield, 1 lb of ethylene produces 2.3 lb ethyl chloride, so the emissions were assumed to be 0.0025 lb ethylene/lb feed and 0.00576 lb ethyl chloride/lb feed. When the reaction product was a solid, the emissions were counted as being entirely composed of the feedstock. The estimated emissions are shown in Table 21-12. The magnitude of emission estimates found by this method are probably within an order of magnitude. However, the identities of the emissions are probably less accurate since a considerable fraction of chemical emissions would probably be reaction by-products, process solvents, catalysts, or other compounds. The fraction of chemical plant effluent which could be burned is hard to estimate, but perhaps from 30 to 60% of the total emissions could be controlled.

Table 21.10 EMISSIONS FROM REFUSE DISPOSAL, MM LBS[a)]

Year	HC	Aldehydes	Organic Acids
1968	1912	50	1470
1970	1966	54	1486
1980	628[b)]	46	206
1990	954	68	308
2000	1444	100	464

a) Totals for municipal and industrial incineration.

b) Assume pollution control legislation will eliminate open burning by 1980.

Table 21.11 EMISSIONS FROM CARBON BLACK MANUFACTURE, MM LBS

Year	Production, MM lbs	HC
1968	2,812	1,574
1970	2,935	1,449
1980	4,700	1,392
1990	6,822	1,627
2000	10,170	2,153

Table 21.12 EMISSIONS FROM PETROCHEMICAL MANUFACTURE, MM LBS

	1968	1970	1980	1990	2000
Non-aromatic Hydrocarbons					
acetylene	2.1	1.9	1.2	0.7	0.4
butylenes	14.4	17.0	31.9	59.7	78.0
cy-hexane	3.0	3.4	8.6	21.4	53.6
ethylene	33.5	41.6	139.9	376.5	1,050.2
higher olefins	6.9	7.9	7.9	7.9	7.9
propylene	14.0	17.3	43.3	108.7	270.3
Total	73.9	89.1	232.8	574.9	1,460.4
Aromatic Hydrocarbons					
benzene	16.3	19.1	42.6	95.6	222.1
styrene	6.5	6.5	19.9	51.4	126.1
toluene	6.0	6.4	9.5	14.0	19.9
xylenes	3.2	2.6	4.2	6.8	11.2
Total	32.0	34.6	76.2	167.8	379.6
Organic Halides					
allyl chloride	0.6	0.7	0.9	1.3	1.9
benzylchloride	0.1	0.1	0.1	0.2	0.3
dichlorobenzenes	0.2	0.2	0.3	0.5	0.7
ethyl chlorides	13.7	13.8	67.6	193.7	560.2
methyl halides	0.7	0.9	1.8	4.3	10.1
vinyl chloride	1.3	1.2	0.7	0.5	0.3
Total	16.6	16.9	71.4	200.5	573.5
Organic Acids and Anhydrides					
acetic acid	0.1	0.2	0.3	0.8	1.8
maleic anhydride	0.5	0.5	1.3	3.3	6.7
phenol	3.1	3.8	9.4	22.7	58.7
xylene base acids	3.2	2.6	4.2	6.8	11.2
Total	6.9	7.1	15.2	33.6	78.4
Alcohols					
butyl alcohols	0.8	1.0	1.8	3.4	4.5
ethyl alcohol	2.6	2.8	3.8	5.3	7.3
isooctyl alcohol	0.6	0.7	1.4	2.6	3.4
isopropyl alcohol	3.8	4.0	6.0	8.9	13.6
methyl alcohol	8.0	10.4	21.1	50.6	118.0
Total	15.8	18.9	34.1	70.8	146.8

Table 21.12 (CONTINUED)

	1968	1970	1980	1990	2000
Aldehydes					
acetaldehyde	1.2	1.4	3.3	7.7	18.1
acrolein	0.1	0.1	0.3	0.7	1.5
butyraldehyde	1.5	1.6	3.5	7.7	16.9
formaldehyde	2.6	3.3	6.7	16.1	37.5
Total	5.4	6.4	13.8	32.2	74.0
Nitrogen-organics					
acrylonitrile	3.7	4.6	9.7	21.8	67.1
aniline	0.5	0.5	0.9	1.6	2.5
methyl amines	0.3	0.3	0.7	1.6	3.7
toluene diisocyanate	0.5	0.5	0.8	1.1	1.6
Total	5.0	5.9	12.1	26.1	74.9
Esters, Ethers, Oxides					
acrylates	1.3	1.6	2.4	5.3	12.0
ethylene oxide	7.1	8.8	20.0	45.4	101.5
glycol ethers	0.3	0.4	0.9	2.1	4.8
propylene oxide	2.1	2.8	10.6	28.2	79.8
Total	10.8	13.6	23.9	81.0	198.1
Ketones					
acetone	1.2	1.5	4.5	12.7	35.2
Total Petrochemical	167.6	194.0	484.0	1,199.6	3,020.9

21.6.7 Emissions from Charcoal Manufacture

Production data for hardwood charcoal manufacture is reported in the 1967 Census of Manufactures.[21-15] It was assumed that hardwood made up 75% of the total charcoal production. Charcoal manufacture emissions are sharply reduced when chemical recovery of the effluents is practiced.[21-1] Since about 10 million lbs/year of methanol are produced from natural sources, while the potential yield from charcoal manufacture is about 80 million lbs/year, it was assumed that chemical recovery is practiced with 12% of charcoal production.21-16) The estimated emissions from charcoal manufacture are shown in Table 19-13. The emission factors were taken from Reference 21-1. Probably almost all combustible emissions from charcoal manufacture could be burned, but it would probably prove cheaper to recover the chemicals as byproducts.

21.6.8 Emissions from Other Sources

Emissions from metallurgical coke manufacture are shown in Table 21-14. The emission factor is taken from Reference 21-1. The potential for afterburner control here is limited, because about 55% of the emissions are associated with loading and unloading the coke ovens. This would present a problem in collecting emissions. Perhaps 20-40% of the emissions could be incinerated.

Emissions from coffee roasting are shown in Table 21-15. Coffee roasters are very common users of afterburners for controlling the malodorous, smoky emissions. The emission factor was taken from Reference 21-1. Coffee roasting emissions are readily incinerated and probably 95% could be treated.

Estimated smokehouse emissions are shown in Table 21-16. Smokehouse emissions are also frequently controlled with afterburners. These emission factors were taken from Reference 21-1. The 1967 Census of Manufactures reports smoked meat production data only for those firms whose primary products are smoked meats.21-15) For this study it was assumed that an additional 25% was produced by wholesalers and retailers. Smokehouse emissions could probably be controlled at the 95% level.

Afterburners have been applied to several metal recovery operations which involve burning off nonmetallic components from the original products. Examples are auto body incineration, brake shoes reclaimation, steel drum reconditioning, and copper wire insulation burnoff. Of these processes, only auto body incineration is a significant emission source. The emissions shown in Table 21-17 for auto body incineration assume that all cars scrapped in a given year are burned and that an average of 7.1% of the car population is scrapped each year.21-27) The estimated automobile population was taken from Reference 21-4, and the emission factors from Reference 21-1.

Emissions from the burn-off step of steel drum reconditioning were estimated by assuming that the emission factors for auto body incineration could be applied to steel drums. This is a very rough approximation, but it should give a result correct within an order of magnitude. Thus it was assumed that incineration of 60 48 lb steel drums produced emissions equal to those from one 2800 lb auto body. Presently, about 50 million drums are being reconditioned each year.21-28) It was assumed that this rate would increase at about 1%/year.21-28) The estimated emissions are shown in Table 21-18. Although the

Table 21.13 EMISSIONS FROM CHARCOAL MANUFACTURE, MM LBS[b)]

Year	Production[a)] M tons	Combustible Particulates	HC	Methanol	Acetic Acid	Total
1968	536	189	47	72	109	417
70	548	193	48	73	112	426
80	835	294	73	112	170	649
90	1273	448	112	170	259	989
2000	1939	683	171	259	396	1509

a) Assuming 75% derived from hardwood and 25% from softwood.

b) Assuming 12% of production associated with chemical recovery since natural methanol production is about 10 MM lbs/yr.

Table 21.14 EMISSIONS FROM METALLURGICAL COKE MANUFACTURE, MM LBS

Year	Production MM tons	HC
1968	62.9	389
70	65	404
80	65	404
90	65	404
2000	65	404

Table 21.15 EMISSIONS FROM COFFEE ROASTING, MM LBS

Year	Coffee Production, MM lb	Organic Acid Emissions	Aldehyde Emissions
1968	2002	1.1	0.2
1970	2043	1.1	0.2
1980	2265	1.2	0.3
1990	2535	1.4	0.3
2000	2794	1.5	0.3

Table 21.16 EMISSIONS FROM SMOKE HOUSES, MM LBS

Year	Total Meat Smoked, MM lbs	HC	Aldehydes	Organic Acids
1968	12,525	0.4	0.5	1.2
1970	12,800	0.4	0.5	1.2
1980	14,169	0.5	0.6	1.3
1990	15,863	0.5	0.6	1.5
2000	17,482	0.6	0.7	1.7

Table 21.17 EMISSIONS FROM AUTO BODY INCINERATION, MM LBS

Year	Cars Scraped, MM	HC	Aldehydes	Organic Acids	Total[a)]
1968	5.68	2.8	1.1	1.7	5.7
1970	5.91	3.0	1.2	1.8	5.9
1980	8.46	4.2	1.7	2.5	8.5
1990	11.6	5.8	2.3	3.5	11.6
2000	13.3	6.7	2.7	4.0	13.3

a) Rows may not add because of rounding.

Table 21.18 EMISSIONS FROM STEEL DRUM RECONDITIONING, MM POUNDS

Year	No. of bbls, MM	HC	Aldehydes	Organic Acids	Total
1968	50	0.42	0.16	0.26	0.84
1970	51	0.42	0.16	0.26	0.84
1980	56	0.46	0.18	0.28	0.92
1990	62	0.52	0.20	0.32	1.04
2000	69	0.48	0.22	0.34	1.14

emission levels are quite low, this process can create a severe local smoke and odor nuisance which can be controlled by afterburning.[21-29)]

Brake shoe debonding emissions are estimated using emission rate data from Reference 21-29. It was assumed that a set of four brake shoes were reclaimed for every 30,000 vehicle miles. Estimated annual mileages were obtained from EPA thru 1990, and these mileages were extrapolated to year 2000.[21-2)] The projected emissions are shown in Table 21-19.

Data on the annual rate at which copper wire insulation is burned off were not found. However, an upper limit for 1970 emissions was obtained by assuming that all the wire and light and heavy copper reclaimed by secondary smelters was burned.[21-30)] Using emission rate data reported in Reference 21-29 for automobile starter field coils and armatures (assuming 50%w copper), it was estimated that the total combustible emissions are less than 2 million lbs per year. Probably 95%+ of the emissions from these reclaimation sources could be controlled with properly designed ovens and afterburners.

Emissions from two additional sources were estimated to provide a useful perspective on the national significance of the combustible emissions problem. Automobile emissions are shown in Table 21-20. The data for hydrocarbons were obtained from EPA, except for the year 2000 estimates, which were extrapolated from the EPA data.[21-2)] The aldehyde and organic acid emissions were estimated using the EPA mileage estimates and emission factors from Reference 21-1. These emission factors do not project any decrease in aldehyde or organic acid emission rates which might result from control devices installed on future automobiles to enable them to meet federally specified carbon monoxide and hydrocarbon emission levels. The federal regulation of automobile emissions should result in a sharp decrease in the relative importance of the automobile as a pollutant source.

Projected emission levels from fuel oil combustion are shown in Table 21-21. They were estimated by using a rough average of the emission factors for various types of fuel oil furnaces given in Reference 21-1. Historical data indicate an approximate 3%/year growth of fuel oil consumption.

This completes the list of processes for which emission rates were projected. The processes studied do not represent an exhaustive list of all sources amenable to afterburning. However, they were chosen to include the important sources in terms of national tonnage, and thus to give a picture of the potential for future air quality improvements through more extensive afterburning of combustible emissions.

Table 21.19 BRAKE SHOE DEBONDING EMISSIONS, MM LBS

Year	No. of Shoes, Millions	HC	Aldehydes	Organic Acids
1968	132	0.02	0.05	0.10
1970	142	0.03	0.05	0.11
1980	192	0.03	0.07	0.15
1990	235	0.04	0.08	0.19
2000	267	0.05	0.10	0.21

Table 21.20 AUTOMOBILE EMISSIONS, MM POUNDS

Year	Vehicle Miles, Billions	HC a)	Aldehydes b)	Organic Acids
1968	990	35,240	762	284
1970	1,068	32,060	848	306
1980	1,440	7,260	1,142	412
1990	1,760	2,420	1,398	504
2000	2,000	2,740	1,582	574

a) Except for the year 2000, the HC estimates are provided by Mr. J. H. Southerland, EPA. The year 2000 data are obtained by extrapolating the EPA estimate for 1990.

b) Aldehyde and organic emissions are based upon the emission factors given by McGraw and Duprey, and assume no controls.

Table 21.21 EMISSION FROM FUEL OIL COMBUSTION, MM LBS

Year	Fuel Oil Consumption, MM bbl	HC	Aldehydes
1968	1543	227	97
1970	1732	255	109
1980	2100	309	132
1990	2700	397	170
2000	4400	647	277

Chapter 22. RESEARCH RECOMMENDATIONS

The objective of this section is to recommend specific research and development plans to provide technology where deficiencies exist, to generate the data required to predict and improve process performance and economics and extend the application of afterburners to additional air pollution sources. Each recommendation contains a description of the purpose, scope, approach and an estimated time and cost schedule. A priority ratio has been assigned to each recommendation based on importance. Two research and development programs are outlined. Plan A is based on an expenditure of one million dollars and Plan B, three million dollars, both for five-year periods. Some projects are common to Plans A and B and some are not. Summaries of programs are contained in Table 22-1 (Plan A) and Table 22-2 (Plan B). The research program is split about 85% on thermal afterburners and 15% on catalytic units. This reflects the more complex engineering questions in thermal units. The levels of fundamental studies on the thermal and catalytic units are about the same.

22.1 Thermal Afterburners

The research and development program is aimed at the combustor with fundamental and engineering studies included. Research effort on auxiliary equipment like heat exchangers and fans is not considered necessary.

22.1.1 Field Test Data

Plan A

The objective of this project is to select appropriate instrumentation and to obtain field data on 10 or so representative types of thermal afterburners. These data would include careful measurements of efficiency as a function of operating conditions on various feeds and types of equipment. This data will provide a basis for correlating efficiency with design parameters and identifying areas for equipment improvement. The system study showed that there was a marked lack of quantitative performance data.

The first part of this project is the selection of simple, low cost, rugged analytical instrumentation that can be used for field monitoring of afterburner performance. This need is common to thermal and catalytic afterburners and is perhaps more critical for catalytic units since performance monitoring is more important because of catalyst deactivation. The species to be measured are hydrocarbons (typical of fuel streams), partially oxygenated combustion products, carbon monoxide, oxygen, carbon dioxide and particulates. Direct read out of results is highly desirable since afterburner feed conditions are often quite variable. This indicates instrumental methods rather than wet chemical methods. Two sets of instruments to monitor input and output simultaneously would be a marked improvement. A great variety of analytical instrumentation is becoming available and components should be selected from them. A system of instruments should then be field tested for ruggedness, stability and interferences.

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The second part of this project is the collection of field data on representative units (10 or so are envisaged). The units tested should be representative of the various equipment types and the various waste streams burned. Particular attention should be paid to measuring as many of the operating and performance variables as possible. The primary operating variables are flow rate, temperature and residence time. The performance variables are feed concentration and exhaust concentrations with particular attention paid to partially oxygenated combustion products. These data will show in a quantitative way the merits of the various afterburner design features. These data will also provide a basis for correlation of performance on an empirical basis. This project would require 18 months to execute and would cost about $130 M and has a priority 1.

Plan B

With the extended level of effort in Plan B, five additional units would be tested and measurements of temperature profiles and velocity distributions in related units would be undertaken. The additional measurements of temperatures and velocities will permit a better correlation of the field data. The extended program would require a total of 18 months to execute and would cost a total of $175 M and has a priority of 1.

22.1.2 Kinetics of Combustion

Plan A

The objectives of the kinetic study are to measure chemical decomposition rates of the typical waste materials being burnt. These experiments can be made on a laboratory scale in which mixing can be removed as a rate limitation. The chemical decomposition rates are fundamental parameters for a rational design procedure for afterburners. As the system study showed, the controlling rate in many cases is mixing rate in the combustor rather than the chemical oxidation step. However, there are some cases where the chemical rate is limiting, and in most cases mixing and oxidation rates can be made the same order. Furthermore, operating temperatures could be reduced in many cases where the chemical rate is currently many times faster than the mixing rates. The reduced operating temperature will result in fuel economy, increased equipment life and less chance of overheating damage due to feed surges. Measurements of oxidation rate should be made in a laboratory apparatus in which conditions are carefully controlled and measured. Mixing rates of contaminant with hot flue gas can be made very fast in the laboratory so that chemical oxidation rates can be measured as a function of temperature. Much of the literature data now available has mixing and oxidation rates combined. The feeds used should be the hydrocarbons representative of industrial emissions. Rate expressions should be developed relating oxidation rate to hydrocarbon and oxidant concentrations. Correlation methods should be developed to extend the data to a broader range of fuel components. This project would need about one year for execution at a cost of about $110 M with a priority 1.

Plan B

With the extended level of effort in Plan B, this kinetics study should be expanded to include more hydrocarbons, odorous compounds, liquid droplets, solid particulates and materials containing heteroatoms like S, Cl, P, etc. The odorous compounds have the special problem of often being in low concentrations to begin with and requiring high conversions to effect odor nuisance abatement. The liquid droplets have a vaporization step which must occur first which may require additional residence time. If complete vaporization can occur without coking then the rates may not be much lower than for vaporized feeds. If coking occurs or solid particulates are present in the feed then the rates of oxidation of these particules will have to be measured. It is expected that the particulate oxidation rates would be low so that significant increases in residence time would be needed. The oxidation rates of feeds containing heteroatoms like Cl, for example, are expected to be quite low so that kinetic data will be very important in a design. The combustion products must also be measured carefully since they will effect the downstream equipment requirements. The additional work on kinetic studies would take about three years to execute at a most of $290 M. The kinetics of decomposition of hydrocarbons odors and particulates has a priority 1 and heteroatoms a priority 3. The total kinetic study would take about four years at a total cost of $400 M.

22.1.3 Research Afterburner - Design, Construction and Operation

Plans A and B

Design procedures and equipment improvements cannot be developed until engineering experiments are performed on a commercial scale piece of equipment. These experiments will provide the data on efficiency as a function of the measured fundamental parameters, reaction rate, temperature profiles, velocity profiles and mixing rates.

The purpose of this project is to design and build a versatile experimental afterburner. Provision for measurement of the internal variables like temperature, velocity and concentration profiles is required. The ability to change burner arrangements and combustion chamber geometry is to be incorporated into the design. The size of the unit should be selected so that internal conditions are representative of commercial equipment (a size in the range of 2 - 5 MM Btu firing rate would appear reasonable). The design of the unit should be oriented towards convenience in obtaining the required data. Operation of the unit without contaminant feed should be made first and data obtained on temperature and velocity profiles. Also careful measurement of the mixing rates should be made. This provides the fundamental fluid mechanical information necessary for the analysis of the performance efficiency experiments described in subsequent sections. This project will require about one year for execution and will cost about $250 M (includes $75 M capital experditure) and has a priority 1.

22.1.3.1 Experimental Study Hydrocarbon Feeds - Research Afterburner

Plan A

Experiments would be run with various hydrocarbon feeds over a range of operating conditions in the research afterburner described in Section 22.1.3. Operating conditions to be investigated should be feed concentrations, feed temperature, oxygen content, afterburner temperature, and feed flow rate. The analytical techniques developed in Section 22.1.1 would be utilized. Particular attention should be paid to measurement of concentration profiles throughout the afterburner. This project should require about six months to execute for a limited range of hydrocarbon at a cost of $70 M with a priority 1.

Plan B

This would be an extension of the number of hydrocarbons used to a broad range and would require about six months to execute and a total cost about $88 M (priority 1).

22.1.3.2 Experimental Study Odorous Feeds - Research Afterburner

Plan A

This is a similar study to that described in 22.1.3.1 on a range of odorous materials. The special features of odorous materials are that the feed concentrations may be quite low and the required conversion to eliminate odor may be quite high. Characterization of the concentration levels of specific odorous materials that constitute a nuisance should be developed. Additional analytical techniques may be needed. This project should require about six months to execute for a limited number of odorous feeds at a cost of about $70 M (priority 1).

Plan B

This would be an extension of the number of odorous feeds investigated and should require six months of effort at a total cost of $87 M (Priority 2).

22.1.3.3 Experimental Study Particulate Feeds - Research Afterburner

Plan A

A method of adding liquid sprays and solid particulates should be added to the research afterburner described in Section 22.1.3. Analytical apparatus for sampling and particle size analysis should be added to the experimental equipment. A limited range of feed materials would be injected into the afterburner feed at various levels of dispersion. The volatility should range from easily vaporized through low volatility heavy liquids up to non-volatile (but combustile) solids. This project will require six months to execute and will cost about $70 M (priority 2).

Plan B

This will be an extension to a broad range of feed compounds and will require six months at a total cost of $87 M (priority 2).

22.1.3.4 Experimental Study of Oil Firing - Research Afterburner

Plan B

The previous experimental work was all based on gas firing as the supplemental fuel. This is the supplementary fuel that is commonly used in existing afterburners, however, natural gas supplies are becoming limited and the cost of natural gas will increase. Therefore, in the future, there will be increasing use of oil firing in afterburners. A gun for oil firing will be added to the research afterburner described in Section 22.1.3. Experiments will then be conducted with representative feeds from the hydrocarbons odors and particulates to determine the difference between oil and gas firing. This project could be executed in a period of one year at about a cost of $187 M (priority 3).

22.1.3.5 Experimental Study Heteroatom Containing Feeds - Research Afterburner

Plan B

The afterburner effeciency will be measured for effluent streams containing heteroatoms like chlorine, sulfur, phosphorous, which result in flame inhibition, increased residence time requirements, corrosive conditions and requirement for stack gas scrubbing. The research afterburner will need additional equipment for stack gas scrubbing and for corrosion protection in the combustor. The experiments would cover a range of feeds containing heteroatoms. The operating problems will be quite difficult with these feeds so that a period of about two years would be required at a cost of about $350 M (including $50 M capital) with a priority 3.

2.1.4 Research Afterburner - Gun and Control System Development

Plan A

A common problem in operation of afterburners is the variations in feed rate and composition occurring in the process generating the effluents. This causes instability in the burner operation, overheating of the equipment and erratic performance. New burner designs and integrated control systems are needed to obtain a much better turndown ratio than is currently achievable. An experimental attack on this problem is proposed using the research afterburner described in Section 22.1.3. This project would require about one year for execution and would cost about $75 M (includes 25 M capital) and is priority 1.

Plan B

This would be an expanded version of the above project with the objective of achieving a larger turndown ratio. This project would require one year for execution and would cost a total of $150 M (includes $25 M capital) and is priority 1.

22.1.5 Mathematical Modelling of Afterburners

Plan A

The kinetic data obtained in Section 21.1.2 and the engineering data obtained in Sections 21.1.3.1, 21.1.3.2, and 22.1.3.3 will be a reliable base upon which to build a mathematical model. The model should include the interaction of the chemical kinetics, and the physical processes of mixing and heat transfer. The measured values of concentration, temperature and velocity profiles will provide a proper test of the validity of the model. Existing methods of solving the fluid mechanical problems can be utilized. A successful model development will provide the basis for a design method which can be used to design cheaper, more efficient afterburners. This initial plan will cover hydrocarbon, odor and particulate feeds. This project can be executed in about 2 years at a cost of $50 M and has a priority of 1.

Plan B

Extension of the model to include oil firing and heteroatom containing feeds. This could be executed in four years at a total cost of $400 M and has a priority of 1.

22.1.6 Field Implementation of New Designs

Plan B

Utilizing the design methods developed in Section 22.1.5 prepare designs for two afterburners utilizing new ideas developed in the earlier work. These units should have a lower cost, higher efficiency than units currently in use and would hopefully be close to optimum design. These units should then be constructed and installed in a process environment. This project would cost about $250 M (include $100 M capital) and would require about one year for execution (priority of 3).

22.1.7 Field Test New Afterburners

Plan B

Field tests on the two new afterburners will determine the performance under realistic industrial conditions. This will demonstrate the applicability of the new technology developed in this overall research program. This project would cost about $150 M and could be executed in about one year and has a priority 3.

22.2 Catalytic Afterburners

The research program here focusses on the fundamental kinetics since it has been shown in the handbook that the engineering design methods are available already.

22.2.1 Field Test Data

Plan A and B

The systems study showed a marked lack of quantitative performance data so the objective of this project is to obtain this data using the analytical techniques of Section 22.1.1. The good and bad features of the operating catalytic afterburners can then be properly assessed with the data. Correlation and analysis of this performance data can be made on the basis of the design methods discussed in the handbook. This project could be executed in one year at a cost of about $75 M and has priority 1.

22.2.2 Kinetics of Catalytic Oxidation

Plan A

The objective is to measure the chemical rate of oxidation for typical catalyst and waste streams. The handbook shows that mass transfer and chemical kinetics interact closely in the performance of catalytic units. A more rational design can be made if the fundamental kinetic rate of oxidation is known. A laboratory reactor should be built to measure the rates of oxidation over selected catalysts commercially available for selected feed materials. The laboratory reactor must be designed such that the chemical rate is limiting rather than the mass transfer rate. Kinetic correlations should be developed relating the oxidation rate to contaminant concentration, oxygen concentration and temperature. This project could be executed in one year at a cost of $50 M and has a priority 1.

Plan B

Extension to a wider range of commercial catalysts and feed materials. Execution in one year at a total cost of $100 M with a priority of 1.

22.2.3 Catalyst Poisoning and Deactivation Rates

Plan A

Catalyst life is an important factor in catalytic afterburner costs. A laboratory reactor will be built for purposes of measuring catalyst life. This should be an automated, unattended apparatus so that long run times can conveniently be obtained. These life tests would use feeds studied in Section 22.2.2. In addition, the poisoning rates of trace materials commonly found in effluent streams should also be determined for commercially available catalysts. While these trace materials may not constitute an effluent requiring control they will have to be removed prior to catalytic oxidation if they poison the catalyst. This project could be executed in about one year at a cost of $50 M and has a priority of 1.

Plan B

Extend range of possible catalyst poisons, feed streams and commercial catalysts. Execute in one year at a total cost of $100 M with a priority 1.

22.2.4 New Catalyst Development

Plan B

The cost of the noble metal catalysts is a significant portion of the total cost of a catalytic unit. Noble metal catalysts are also poisoned rapidly by certain contaminants. The objective of this project is to find base metal catalysts that are lower cost and are more resistant to poisoning. These measurements should be made in the laboratory apparatus described in Section 22.2.2. The research effort in automobile exhaust catalysts should provide some catalysts that may be applicable to stationary afterburners. This project could be executed in one year at a cost of about $150 M and has a priority 2.

Table 22-1. RESEARCH PROGRAM, PLAN A, $1 MM

Project		Cost $ M	Year					Priority
			1	2	3	4	5	
Thermal								
22.1	Field test data	130	—	→				1
22.1.2	Kinetics of combustion	110		→				1
22.1.3	Research afterburner - Design, construction and operation	250			→			1
22.1.3.1	Experiments, hydrocarbon feeds	70				→		1
22.1.3.2	Experiments, odorous feeds	70				→		2
22.1.3.3	Experiments, particulate feeds	70					→	2
22.1.4	Research afterburner - Gun and control system development	75				→		1
22.1.5	Mathematical modelling of afterburner	50				—	→	1
Catalytic								
22.2.1	Field test data	75	—	→				1
22.2.2	Kinetics of catalytic oxidation	50	→					1
22.2.3	Catalyst poisoning and deactivation rates	50		→				1
Cost per year			180	235	250	240	95	
Total cost		1000						

Table 22-2. RESEARCH PROGRAM, PLAN B, $3 MM

Project		Cost $ M	Year 1	2	3	4	5	Priority
Thermal								
22.1	Field test data	175	——	→				1
22.1.2	Kinetics of combustion (hydrocarbons, odor particles)	200	——	→				1
	Kinetics of combustion (heteroatoms)	200			——	→		3
22.1.3	Research afterburner - Design, construction and operation	250	→					1
22.1.3.1	Experiments, hydrocarbon feeds	88		→				1
22.1.3.2	Experiments, odorous feeds	88		→				2
22.1.3.3	Experiments, particulate feeds	87			→			2
22.1.3.4	Experiments, oil firing	187			——	→		3
22.1.3.5	Experiments, heteroatoms	350				——	→	3
22.1.4	Research afterburner - gun and control system development	150			→			1
22.1.5	Mathematical modelling of afterburner	400		——	——	——	→	1
22.1.6	Field implementation of new designs	250				→		3
22.1.7	Field test new afterburners	150					→	3
Catalytic								
22.2.1	Field test data	75	——	→				1
22.2.2	Kinetics of catalytic oxidation	100	→					1
22.2.3	Catalyst poisoning and deactivation rates	100		→				1
22.2.4	New catalyst development	150			→			2
Cost per year			613	562	675	750	400	
Total cost		3000						

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Reliable tubes may be purchased from Metronics Associates, Inc., 3201 Porter Drive, Stanford Industrial Park, Palo Alto, California 94304. Office of Standard Reference Materials, National Bureau of Standards, Washington, D. C. 20234 will supply a calibrated and certified SO_2 tube; order SRM 1625. Other sources are Polyscience Company, P. O. Box 65, Kenilworth, Illinois 60043 and Analytical Instrument Development Company, West Chester, Pennsylvania 19380.

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